COMPARISON OF BRINE PRODUCTION METHODS AND CONVERSION PROCESSES FOR GEOTHERMAL ELECTRIC POWER GENERATION

by

DAVID G. ELLIOTT

EQL REPORT No. 10
July 1975

Environmental Quality Laboratory
CALIFORNIA INSTITUTE OF TECHNOLOGY
Pasadena, California 91125

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GEOTHERMAL RESEARCH STUDY IN THE
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SUMMARY OF GEOTHERMAL RESEARCH STUDY
IN THE SALTON SEA REGION OF CALIFORNIA

I Introduction

This document is one of several that report the results of a study carried out by the EQL during 1975. This investigation was directed at definition of and support for a geothermal resource demonstration program to be conducted at the Salton Sea KGRA.* The originating agency was NSF/RANN, but in January of 1975 the responsibility for the project was transferred to ERDA. It was anticipated that the results of the research would be continuously fed in to the demonstration program, which was originally anticipated to be initiated during the period of this present grant. However, the demonstration program was not initiated during that time. Those programs are now getting underway, and it is hoped that these results may be useful to ERDA and the contractors in arriving at a program of research. A summary of results in each of the several research areas is given in the following paragraphs, in order that the reader can place the present report in context.

II Engineering Studies

For high-temperature reservoirs the technique of pumping wells was found to be inferior to flashing flow (self-flow), as theory suggests a considerable advantage to be gained by increasing well diameter in the upper (flashing) parts of the well bore. A more thorough analytical treatment, with experimental verification, is recommended. This technique may offer a major savings in field development cost.

For high-temperature reservoirs it was found that multiple-flash direct steam systems offer superior performance as compared to binary processes. Total-flow expanders offer promise, however, and their development is recommended. Even for low-temperature reservoirs, binary processes offer no significant theoretical performance advantage as compared to flash processes. It was not possible to carry out cost comparisons during the study.

*KGRA - Known Geothermal Resource Area, a legal definition.

This work is reported in EQL Report No. 10, "Comparison of Brine Production Methods and Conversion Processes for Geothermal Electric Power Generation."

III Environmental Geochemistry

The review of trace element analyses of Salton Sea KGRA brines indicates that more sophisticated sampling and analytic techniques will have to be applied to be sure that minor constituent content is properly understood. Generally, environmental problems will come from volatile species. A program of analysis is suggested that would indicate what chemical compounds may present environmental hazards. This information is necessary before control requirements could be formulated. The work is reported in an open file EQL document, "Trace Element Problems," available at EQL.

IV Water Chemistry

It was found that much of the existing useful empirical engineering information concerning scale and corrosion is not available in the open literature. Unless this information is released by its owners, a future research program may have to duplicate past experience. Several proposed lines of research on the problem were investigated. Some of the most basic approaches, beginning with fundamental chemistry, were judged to be too lengthy or difficult to yield practical results. A basic chemical engineering approach was identified and recommended. The work is reported in EQL Memorandum No. 14, "Brine Chemistry - Scaling and Corrosion."

V Geology

A geology investigation was to have been provided by faculty consultants from University of California at Riverside, but this investigation has proven to be unnecessary. The current geologic knowledge of the Salton Sea KGRA was recently published in the Ph.D. dissertation of Randall. The UCR group had already formulated a continuing program of geological research at the Salton Sea, and proposed the work to NSF. The proposal has been funded. Therefore, no additional research programs are recommended at this time.

VI Reservoir Mechanics

The selected consultants (Petroleum Engineering group at Stanford University) report that methods appropriate to evaluation of the Salton Sea reservoir are contained in the paper "Pressure Transient Analysis for Geothermal Wells" by Henry J. Ramey, delivered at the Second UN Geothermal Symposium in San Francisco. The exact program of measurements can be formulated only when actual well flow programs are determined. No additional research peculiar to the Salton Sea KGRA was recommended. The pressure transient analysis was strongly recommended for implementation during the demonstration program.

VII Economics

Because the Salton Sea KGRA is largely leased, and because business arrangements have been made between producers, utilities and leaseholders, our general study of geothermal economics can help one to understand some aspects of the economics of the Salton Sea field, but cannot be expected to further the demonstration program. The study was carried out and is published in EQL Report No. 11, "The Economics of the Geothermal Energy Industry, Based on Data Through 1974."

VIII Legal and Institutional

A number of legal and institutional issues are often mentioned as hindrances to geothermal development. This study found that the problems are poorly documented in the literature, and it is not possible on the basis of the published record to recommend specific changes in the law or institutions pertaining to geothermal development.* Therefore, a program of research to identify and properly document the issues is recommended. It was also found that the difficulties facing the geothermal industry in general are not met at the Salton Sea KGRA, because so many of the problems have already been overcome there. Therefore it is concluded that a demonstration program at the Salton Sea will do little to illuminate or solve most legal and institutional issues being met elsewhere. The work is reported in an open file EQL document, "Legal and Institutional Factors."

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*Progress in this regard has been made by the recent publication of a Draft Report on Geothermal Energy and the Law by the University of Southern California Law Center.

Acknowledgments

This study was performed by Dr. David G. Elliott, Member of the Technical Staff, Energy Systems Section, Jet Propulsion Laboratory, under the direction of Dr. Martin Goldsmith, Deputy Director, Environmental Quality Laboratory, Caltech. Dr. Goldsmith selected the conversion processes to be analyzed, provided technical guidance, reviewed the results, and participated in writing the manuscript. Valuable technical suggestions were contributed by Professor Rolf Sabersky, Division of Engineering and Applied Science, Caltech.

PREFACE

Before plunging the reader into Dr. Elliott's detailed analysis of geothermal conversion processes, we feel obliged to indicate why the work was carried out. It is not as though no prior investigations have been made; on the contrary, a number of papers and reports exist. What is lacking, however, is agreement. Most (but not all) investigations have agreed that if wells producing dry steam are found, the use of the steam directly in turbines is the preferred means of developing mechanical or electrical power. Where a geothermal field produces hot water, however, (which may partially flash to steam in the well), this degree of unanimity is not found. As early as 1970 public presentations were made[1] which argued for the superiority of conversion systems which would use the produced geothermal water to heat a secondary working fluid, used in a Rankine cycle (thus binary cycle). Subsequent technical papers[2] analyzed system performance and cost superiority. At the same time, however, analyses were published which concluded that binary processes had little to offer.[3]

While in some of the geothermal literature the superiority of the binary process, particularly for lower temperature resources, is taken as assured, in the four hot water geothermal fields worldwide, where extensive commercial exploitation has begun (Wairakei, Cerro Prieto, Ahuachapan, Otake), only direct steam process is used.

In the face of such a diversity of viewpoints, we felt it important to try to understand why the differences exist. The various published analyses were carried out under varying assumptions, and meaningful intercomparison seemed difficult or impossible to accomplish. Therefore we

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[1] See Statements by Harry W. Falk, Jr. and J. T. Kuwada in Compendium of Papers, Imperial Valley - Salton Sea Area Geothermal Hearing, October 22-23, 1970, Resources Agency, State of California.

[2] For example, "The Vapor-Turbine Cycle for Geothermal Power Generation" by J. Hilbert Anderson in Geothermal Energy, ed. by Kruger and Otte,

[3] See, for example, "Geothermal Power" by Basil Wood in Geothermal Energy, UNESCO, 1973, and "The Applicability of the Binary Cycle" by Russel James, Second UN Symposium on Geothermal Resources, 1975.

resolved to conduct our own thermodynamic analysis. This requires two parts -- first the production of geothermal fluids from the reservoir, and, second, the conversion of the heat energy to mechanical (or electrical) form. While most existing geothermal wells are produced by self-flow, some have been pumped. This latter method, which is well matched to binary processes, is compared with self-flow (flashing) in our analysis. Then a variety of direct and secondary conversion processes are compared. This was accomplished for wells typical of the Salton Sea KGRA, the focus of our research. The same analyses were also carried out for a hypothetical case of equally high reservoir temperature but containing low salinity brine. To understand the relative efficacy of the various production and conversion processes for the more commonly-found lower temperature reservoirs, similar calculations were carried out for one such hypothetical case.

Martin Goldsmith
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I. INTRODUCTION

Electric power generation from geothermal brine requires, first, bringing the hot brine to the surface and then converting the heat to electric power (the term brine refers to the hot liquid which may have a composition ranging from pure water to 30 percent salt solution). The various methods of obtaining brine flow from the well fall in the category of brine production, and the various methods of converting the heat to electric power fall in the category of conversion processes. The latter can also be referred to as cycles, but process is a more accurate term because the brine flows only once through the conversion equipment and does not undergo a cyclic change.

Two brine production methods will be compared with respect to available power at the wellhead: self flowing (brine lifted by steam from vaporization of the brine), and pumped (brine lifted by a mechanical pump and kept in the liquid state). Five conversion processes will be compared with respect to fraction of available power converted to electric power: flash steam (steam turbines operating on steam from flash vaporization of the brine), dual steam (flash steam with two-phase expanders to recover the flash vaporization mechanical energy), total flow (two-phase expanders only), binary (heat transferred from the brine to a secondary working fluid), and flash binary (heat transferred from flashed steam to a secondary working fluid).

II. PRODUCTION METHODS

A. Method of Calculation

The method of calculation for self flowing is derived in Appendix A. Given the reservoir temperature, brine concentration, and well depth, a flow rate is chosen and the well bottom pressure is calculated from the drawdown pressure factor, the decrease in well bottom pressure below reservoir pressure per unit flow rate (a number available from measurement or from reservoir theory).

Next, the location of the flash level where the pressure in the well reaches saturation is calculated taking into account the hydrostatic pressure difference and the friction pressure drop for given well diameter. The flow conditions at 1°C intervals between the flash level and the wellhead are then calculated by balancing pressure force against hydrostatic and friction forces and momentum change. The result is the temperature, pressure, and quality (fraction of total flow in the vapor phase) at the wellhead. The energy available from an isentropic expansion of unit mass of the wellhead product (after stagnation of the flow at constant pressure) to some specified rejection temperature is then calculated as the specific available wellhead power. Finally, multiplying the specific available wellhead power by the flow rate gives the total available wellhead power.

The method of calculation for pumping is derived in Appendix B. The type of pump assumed is one driven by a steam turbine using heat from the brine flowing up the well.[1] Such a pump is characterized by a cooling factor which is the reduction in wellhead temperature per unit of pump pressure rise. The pump pressure rise required to maintain saturated liquid at the wellhead is calculated taking into account the drawdown pressure at the bottom and the hydrostatic and friction pressure drops in the well. The wellhead temperature is then calculated from the cooling factor, the required wellhead pressure is recalculated, and the procedure is repeated until the correct wellhead temperature is found. The specific available wellhead power and total available wellhead power are then calculated.

The thermodynamic properties of pure water and steam are taken from the Steam Tables.[2] The thermodynamic properties of brine and of steam in equilibrium with brine are calculated by the procedure of Appendix C using data for brines in which the dissolved salt consists of KCl, $CaCl_2$, and NaCl in the ratio 1.00:1.95:3.55 by mass, typical of some geothermal brines in the Salton Sea area.[3]

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(1) Matthews, H. B., and McBee, W. D., "Geothermal Down-Well Pumping System," Conference on Research for the Development of Geothermal Energy, NSF-RA-N-74-159, National Science Foundation, Washington, D.C., 1974.

(2) Keenan, J. H., and Keyes, F. G., Steam Tables (Metric Units), Wiley, New York, 1967.

(3) Pool, M. J., and Nevens, T. D., Determination of Thermodynamic Properties of Brines, Report No. 2151, Denver Research Institute, Denver, Colorado, February 1964.

B. Well Constants

The main variables affecting wellhead available power are reservoir temperature and brine concentration. Drawdown pressure factor (at the low values required for practical wells), well-bottom pressure, and well depth have a lesser effect, and well diameter is restricted to a narrow range by available drilling techniques; therefore, fixed values are used for those quantities (except for one example which will be presented of higher drawdown pressure). Table 1 lists the well constants adopted. The depth is 1500 m (most geothermal wells have depths between 1000 and 2000 m). The inside diameter of the casing is 0.25 m, a typical value for production wells. Calculations are also made for a stepped diameter well with the casing diameter doubled to 0.5 m above the flash level.

Reservoir temperatures range from 150°C, about the lowest temperature considered feasible for geothermal electric power generation, to 300°C, about the highest temperature encountered. The reservoir pressure is assumed to be that due to the normal hydrostatic pressure of 20°C ground water at 1500 m depth, namely 14.7 MPa (megapascals) or 2130 psi. The drawdown pressure factor assumed is 25 kPa (3.6 psi) per kg/s, the value measured for a typical well (No. 1 IID in the Salton Sea area).[4] The brine concentration varies from zero (pure water) to 30 percent, the largest usually encountered.

The volume of noncondensable gases in geothermal wells is usually not large enough to affect the flow conditions. In conversion processes the presence of noncondensables affects the ability to achieve vacuum condensing but does not otherwise greatly affect the power output. Therefore, noncondensables are assumed absent for purposes of calculating well flow and power output. The problem of condensing limitations due to noncondensables will be considered separately.

The skin friction coefficient, which is the ratio of wall shear to dynamic pressure (using the two-phase mixture density) is assumed to be 0.008, about twice the value for smooth pipe; this value gives the best agreement with data for a particular well (No. 1 IID) as discussed in Appendix A. The value of slip

(4) Helgeson, H. C., "Geologic and Thermodynamic Characteristics of the Salton Sea Geothermal System" American Journal of Science, Vol. 266, pp. 129-166, March 1968.

Table 1 Well Constants

Depth	1500 m
Diameter (constant diameter well)	0.25 m
Diameter (stepped diameter well)	0.50 m from top to flash level or 750 m, whichever is less, and 0.25 m from there to bottom
Reservoir temperature	150 to 300°C
Reservoir pressure	14.7 MPa (2130 psi)
Drawdown pressure factor	25 kPa (3.6 psi) per kg/s
Brine concentration (Reference-3 composition)	zero and 30%
Noncondensables	zero
Skin friction coefficient	0.008
Interphase slip velocity	zero
Heat transfer out of well	zero
Conversion process rejection temperature	45°C
Pump cooling factor for pumped well	4°C per MPa (5°F per 100 psi)

Table 2 Flash Levels in Stepped Diameter Wells at Peak-Power Flow Rates

Reservoir Temperature °C	Brine Concentration percent	Depth of Flash Level m	Depth of 0.5 m Diameter Casing m	Peak Available Power MW
150	0	300	300	6
200	0	660	660	22
250	0	950	750	43
300	0	1400	750	61
150	30	400	400	2
200	30	640	640	10
250	30	920	750	23
300	30	1270	750	37

velocity between the phases that gives the best agreement with the same data is zero. Heat transfer out of the well is assumed to be zero, a valid approximation at the large flow rates of interest.

The conversion-process rejection temperature for calculating available power is assumed to be 45°C, a typical powerplant condensing temperature. The cooling factor for pumping (the reduction in wellhead temperature per unit of pump pressure rise due to vaporizing the working fluid to drive the pump) is assumed to be 4°C per MPa (5°F per 100 psi). This factor is the lowest projected in Reference 1 and half the factor predicted there for early pumps.

C. Effect of Stepped Diameter

With self flowing, the volume of steam increases rapidly near the top of the well due to the combined effects of increasing quality and decreasing pressure. The upper part of the well acts, therefore, as a throat, and the area at the top of the well is a dominating factor in setting the flow rate. Furthermore, the sonic velocity in two-phase mixtures is relatively low, 100 to 200 m/s at typical wellhead conditions. Consequently, the wellhead flow can easily reach sonic velocity. At that condition the well is said to be choked because any further reduction in pressure outside the top of the well casing will cause no further decrease in pressure inside the casing and no increase in flow rate. Choking is shown both experimentally and theoretically in Figure 1, which presents the flow rate of the No. 1 IID well[4] as a function of wellhead pressure. Based on the calculation method of Appendix A, the wellhead pressure at zero flow is 3.4 MPa (493 psi), and as the wellhead pressure is reduced the flow rate rises to a maximum of 64 kg/s at a wellhead pressure of 0.64 MPa (93 psi). At that condition the velocity of the flow leaving the top of the casing is equal to the sonic velocity of 170 m/s in the two-phase mixture. Any further reduction of pressure (outside the top of the casing) gives no higher flow rate. The measured flow rate (dashed curve) from Figure 10 of Reference 4 shows the same behavior, although with a higher peak flow rate and higher zero-flow pressure than predicted by the theory.

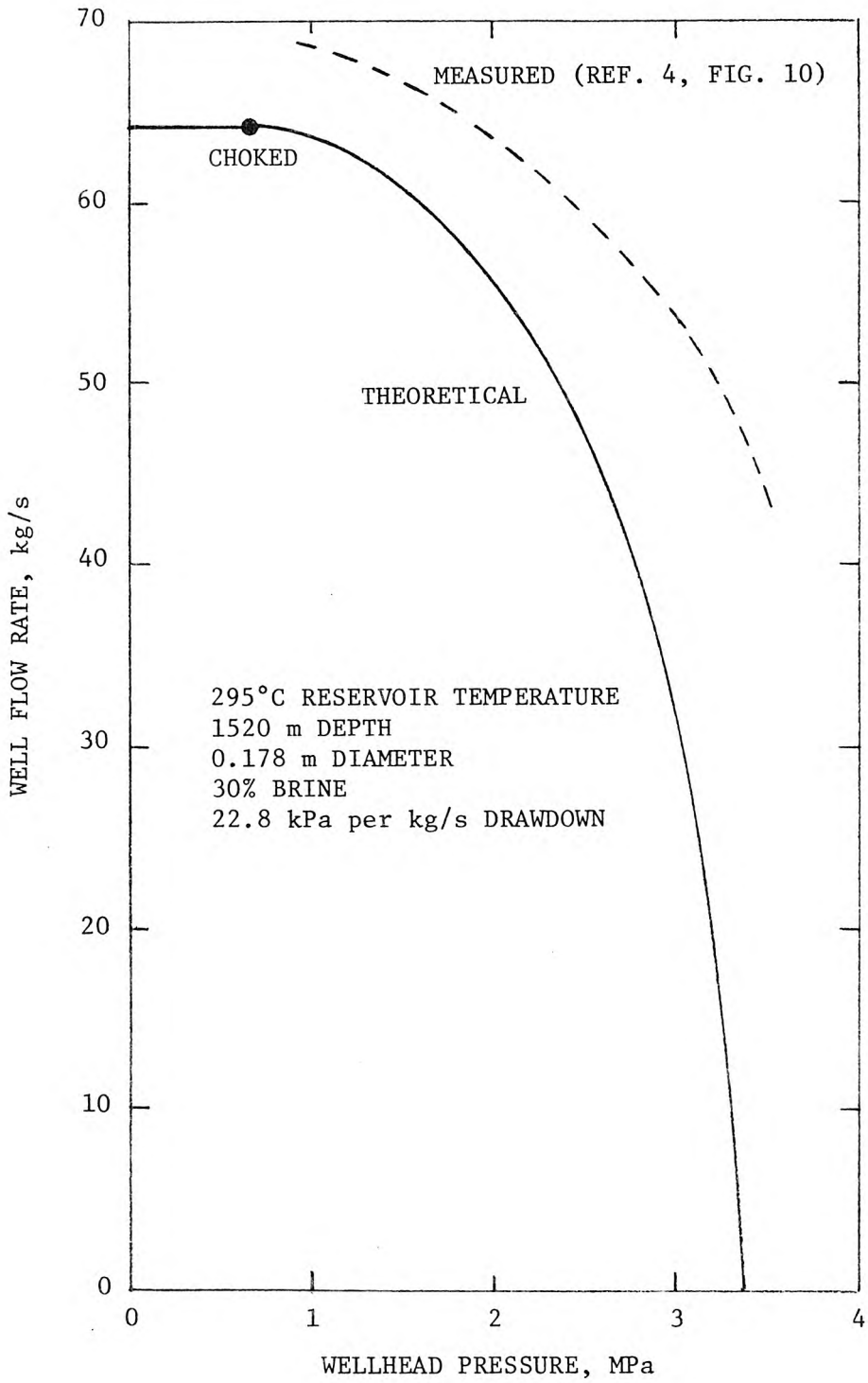

Fig. 1 Comparison of measured and theoretical total well flow rate for No. 1 IID well[4]

The maximum available total power at the wellhead occurs just before choking when the velocity is about half of sonic. Figure 2 shows the variation of Mach number (ratio of flow velocity to sonic velocity) between the wellhead and the flash level for the well constants of Table 1 with 150°C reservoir temperature and zero brine concentration (pure water). For the constant diameter well the flash level is at a depth of 87 m for a flow rate of 55 kg/s, the flow rate giving the highest total available power at the wellhead. The Mach number for this flow rate, shown by the solid curve, rises rapidly to 0.1 in the first 10 m above the flash level, rises more slowly to 0.3 in the next 60 m, and increases rapidly to 0.5 in the last 17 m to the wellhead. If the flow rate is increased another 2 percent to 56 kg/s the Mach number increases to 1 at the wellhead.

If the diameter of the casing is doubled between the wellhead and the flash level a higher flow rate (121 kg/s) can be carried before choking occurs. The total available wellhead power peaks at 116 kg/s with the flash level at 300 m; the variation of Mach number with depth for that condition is shown by the dashed curve in Figure 2.

If the enlarged casing extends only part way to the flash level the effect on total available wellhead power is that shown in Figure 3. The total available wellhead power is 3.2 MW at the peak-power flow rate of 55 kg/s for the constant diameter well. If the well diameter is doubled to 0.5 m between the wellhead and the flash level, the peak-power flow rate is increased to 116 kg/s and the available power is doubled to 6.3 MW. There is no gain from extending the enlarged casing below the flash level of 300 m; and if the enlarged casing does not reach the flash level the power is reduced in proportion.

If the casing diameter is tripled to 0.75 m between the wellhead and the flash level, the upper curve in Figure 3 shows that the peak-power flow rate is increased to 179 kg/s, the flash level is lowered to 500 m, and the available power is increased to 9.3 MW, almost three times the available power with constant diameter.

At higher reservoir temperatures the flash levels are deeper, approaching the bottom of the well at 300°C reservoir temperature. For this reason, the

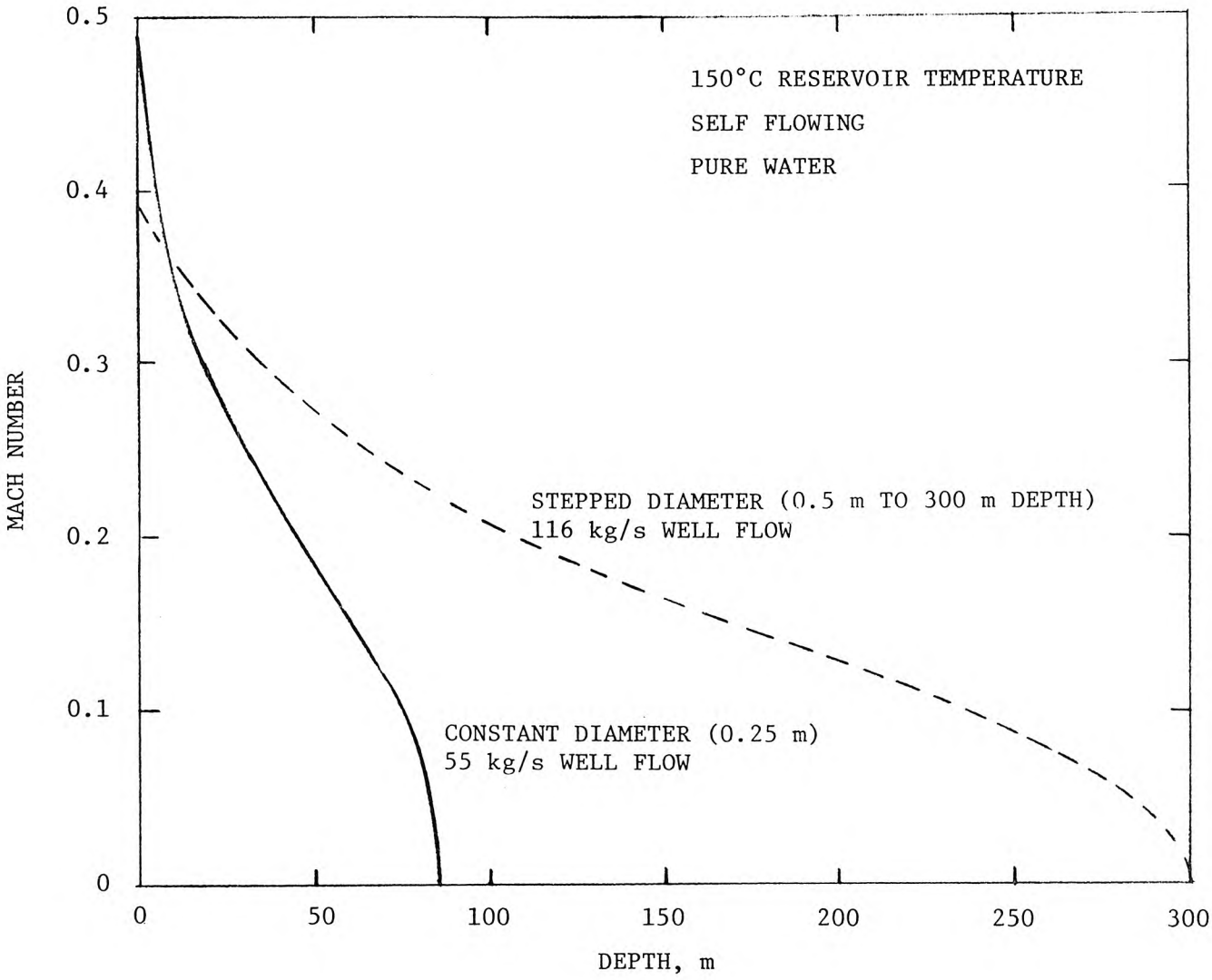

Fig. 2 Variation of Mach number with depth for constant diameter and stepped diameter wells at the peak-power flow rates

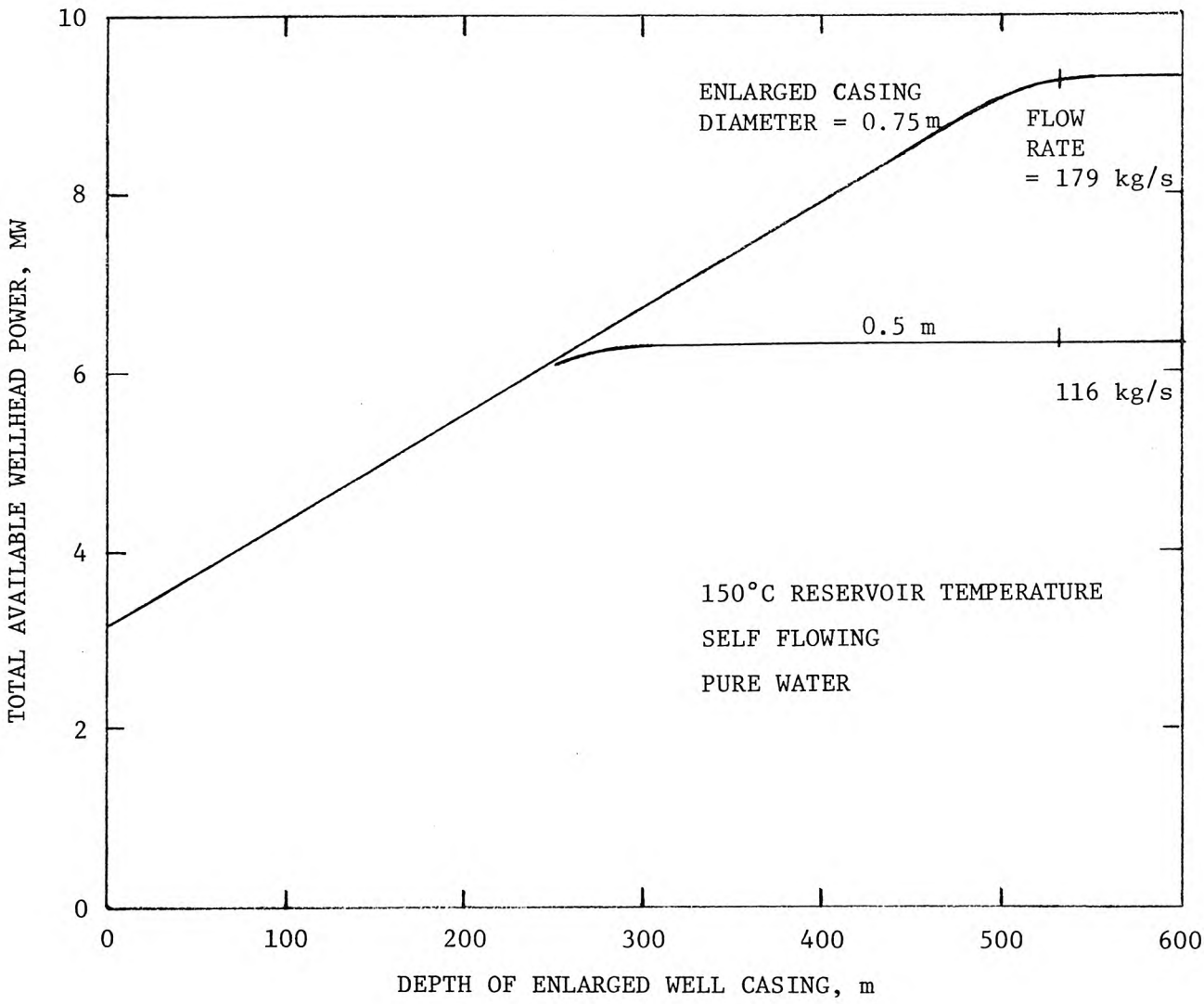

Fig. 3 Effect of enlarged casing diameter and depth of enlarged casing on total available wellhead power for 150°C reservoir temperature

stepped-diameter cases are arbitrarily calculated for enlarged casing extending only to 750 m (half way to the bottom) if the flash level is below that depth. Table 2 (page 4) gives the flash levels at the peak-power flow rates and the enlarged casing depths for the stepped-diameter wells.

D. Effect of Flow Rate on Available Wellhead Power

The specific available wellhead power for 150°C reservoir temperature and zero brine concentration is plotted as a function of flow rate in Figure 4. At flow rates below 10 kg/s the well is self flowing with liquid at the wellhead, the well being pumped by the difference in density between the hot water inside and the cold ground water outside. The specific available wellhead power is 62 J/g (62 kW per kg/s). As the flow rate is increased with self flowing (dashed curves in Figure 4) the specific available wellhead power decreases due to friction and due to unrecovered kinetic energy (assumed lost except for the portion recovered as heat in stagnation of the flow at the wellhead). The constant diameter well chokes at 56 kg/s with the specific available wellhead power reduced to 55 J/g. The stepped-diameter well chokes at 121 kg/s with the specific available wellhead power reduced to 47 J/g.

The specific available wellhead power with pumping is shown by the solid curve in Figure 4. The specific available wellhead power decreases more rapidly with pumping than with self flowing because of the losses in converting the heat of the brine to mechanical pumping power (physically, the power lost is rejected from the condenser on the surface where the steam from the down-well turbine driving the pump is condensed). Thus, the losses in the pump and turbine are greater than the friction and kinetic energy losses with self flowing, until the self-flowing well approaches choking conditions.

As the flow rate with pumping is increased, it is necessary to locate the pump at increasing depths in the well for the pump inlet pressure to remain above saturation. As shown in Figure 4, the limiting case of the pump located at the bottom with inlet pressure equal to saturation pressure occurs at a flow rate of 570 kg/s. The specific available wellhead power at that condition is almost zero

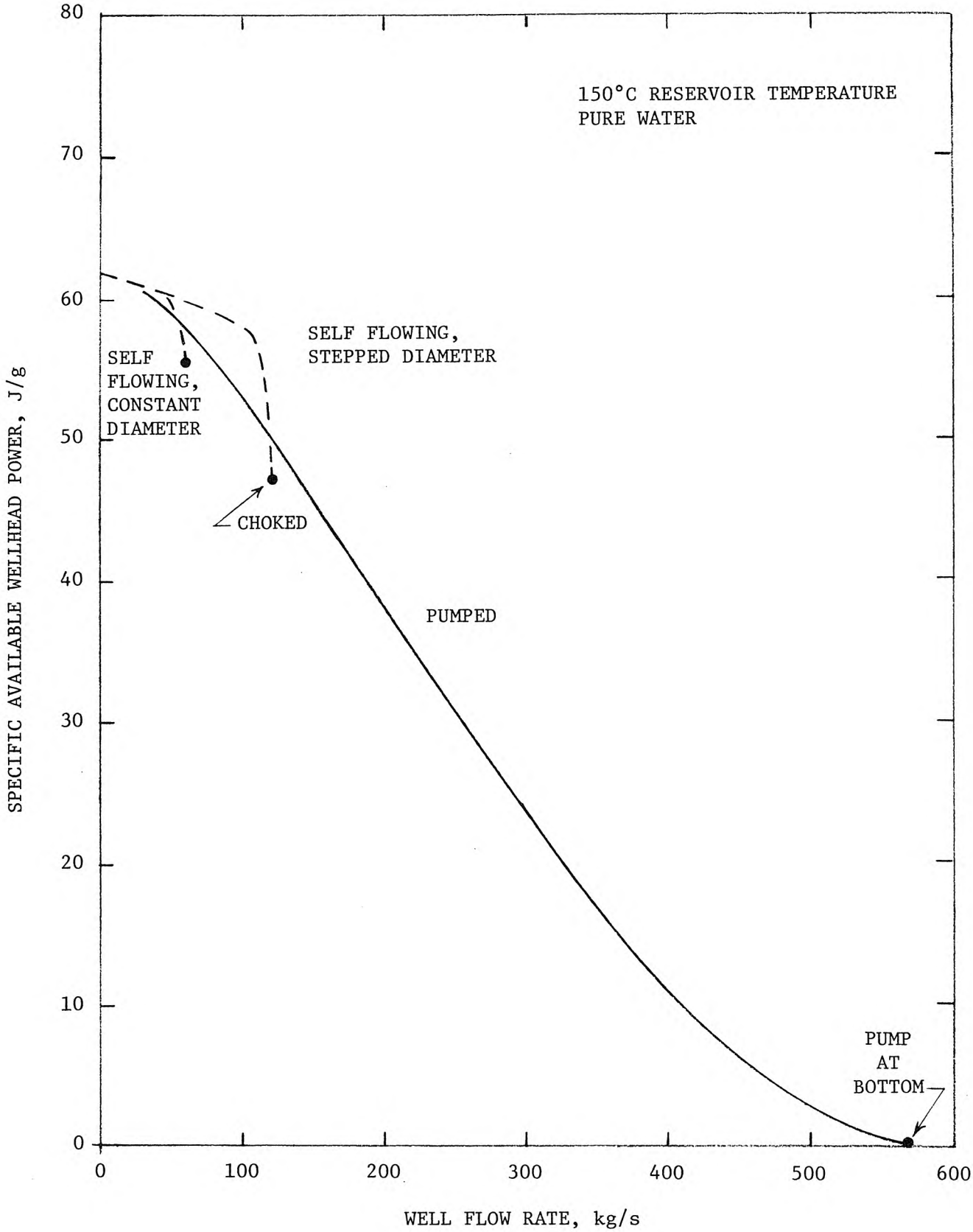

Fig. 4 Comparison of specific available wellhead power with self flowing and pumping for 150°C reservoir temperature and pure water

because the well flow must be cooled almost to 45°C to drive the pump. This is an impractical limiting case shown only for completeness.

Figure 5 presents the total available wellhead powers (the Figure 4 values multiplied by flow rate) as a function of flow rate for 150°C reservoir temperature and zero brine concentration. For self flowing at constant diameter the total available wellhead power reaches a maximum of 3.2 MW at 55 kg/s. The power drops to 3.1 MW when the well is choked at 56 kg/s (the increased flow rate is not enough to compensate for the increased friction and kinetic energy losses). For self flowing with stepped diameter (0.5 m diameter to 300 m depth and 0.25 m diameter from there to the bottom) the total available wellhead power reaches a maximum of 6.3 MW at 116 kg/s and drops to 5.7 MW when the flow is choked at 121 kg/s.

For pumping, the total available wellhead power increases more slowly with flow rate than for self flowing because of the more rapid decrease in specific available wellhead power. The total available wellhead power reaches a maximum of 7.6 MW at a flow rate of 210 kg/s and then decreases with further increase in flow rate because the pump-loop losses increase faster than the flow rate. At 210 kg/s the pumping power (volume flow rate times pump pressure rise), shown by the lower curve in Figure 5, is 1.3 MW; the pump must be at a depth of 600 m for saturation inlet pressure (and at a depth 20 to 100 m lower for practical inlet pressures).

The specific available wellhead power for 300°C reservoir temperature and zero brine concentration is plotted as a function of flow rate in Figure 6. The powers are about five times higher than for 150°C reservoir temperature, but the variation with flow rate is similar. At this temperature, the self-flowing well delivers vapor to the surface at all flow rates. The specific available wellhead power varies from 317 J/g at low flow rate to 250 J/g at choking with constant diameter and to 210 J/g at choking with stepped diameter (0.5 diameter to 750 m depth).

With pumping, a pressure rise of 3 MPa (440 psi) is required at zero flow to provide saturation pressure at the wellhead, and the specific available wellhead

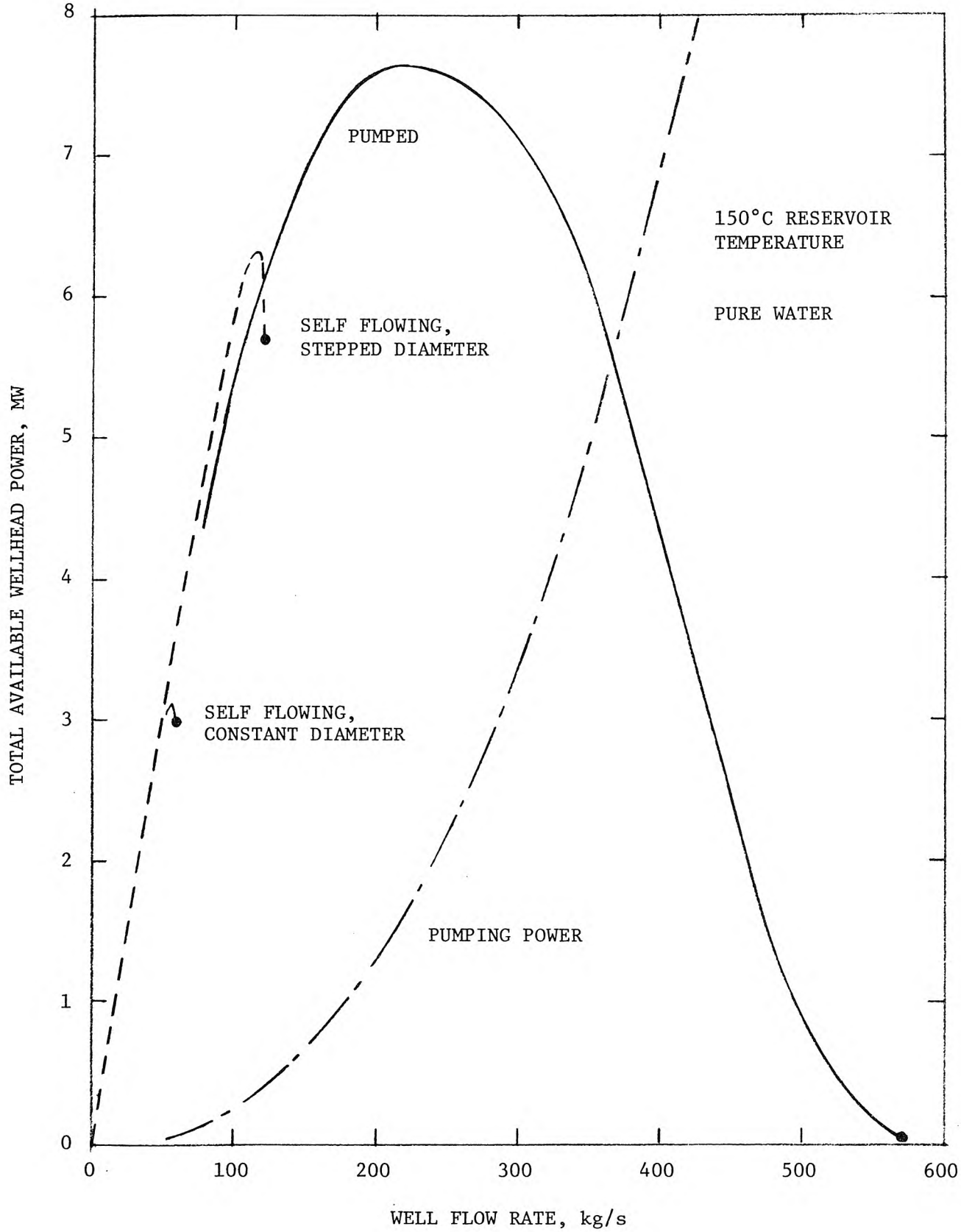

Fig. 5 Comparison of total available wellhead power with self flowing and pumping for 150°C reservoir temperature and pure water

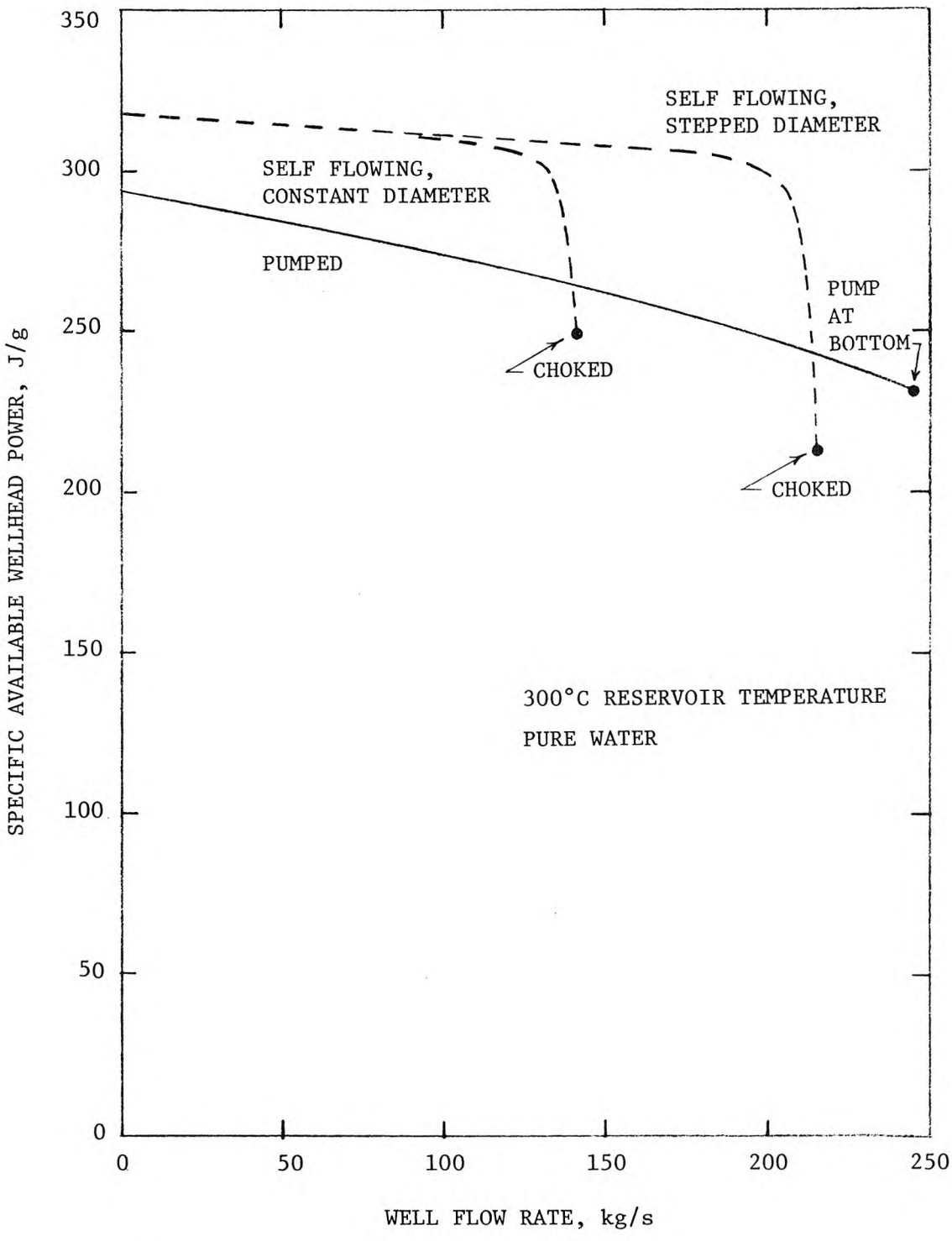

Fig. 6 Comparison of specific available wellhead power with self flowing and pumping for 300°C reservoir temperature and pure water

power is already reduced to 290 J/g. With increasing flow rate the specific available wellhead power decreases more rapidly than with self flowing (except near choking) and drops to 230 J/g in the limiting case with the pump at the bottom at saturation inlet pressure. The absolute decrease in specific available power with flow rate is about the same as at 150°C reservoir temperature, but the relative decrease is less because of the higher available power.

The total available wellhead power with 300°C reservoir temperature and zero brine concentration is presented in Figure 7. With self flowing the constant-diameter well provides 40 MW available power and the stepped-diameter well 61 MW. Pumping gives the same total available wellhead power as self flowing at constant diameter if the pumping power is 1.4 MW and the pump is at 1200 m depth, but even the limiting case of a 3.4 MW pump at the bottom gives less available wellhead power than self flowing with enlarged casing to 750 m.

The results for 30 percent brine concentration are shown in Figures 8-11. The main difference compared with pure water is lower specific available power due to the lower enthalpy of the brine. In addition, the higher density of the brine reduces the peak-power and choking flow rates with self flowing at low reservoir temperature. With 150°C reservoir temperature and 30 percent brine concentration (Figure 9) the total available wellhead power with pumping can be six times as great as for self flowing at constant diameter and almost twice as great as for self flowing with stepped diameter. However, this is not a realistic case since no low temperature high-salinity geothermal resources are known to exist. At 300°C and 30 percent brine (Figure 11) even a 5 MW pump at the bottom of the well gives only twice as much wellhead power as self flowing with constant diameter and negligibly more power than self flowing with stepped diameter.

E. Comparison of Available Wellhead Power for Self Flowing and Pumping

The best choice of flow rate for self flowing is the flow rate that gives maximum total available wellhead power, assuming there are no gains elsewhere from using lower flow rate. The best flow rate for pumping, however, will be at least slightly below the peak-power flow rate because reduced pump power and depth result.

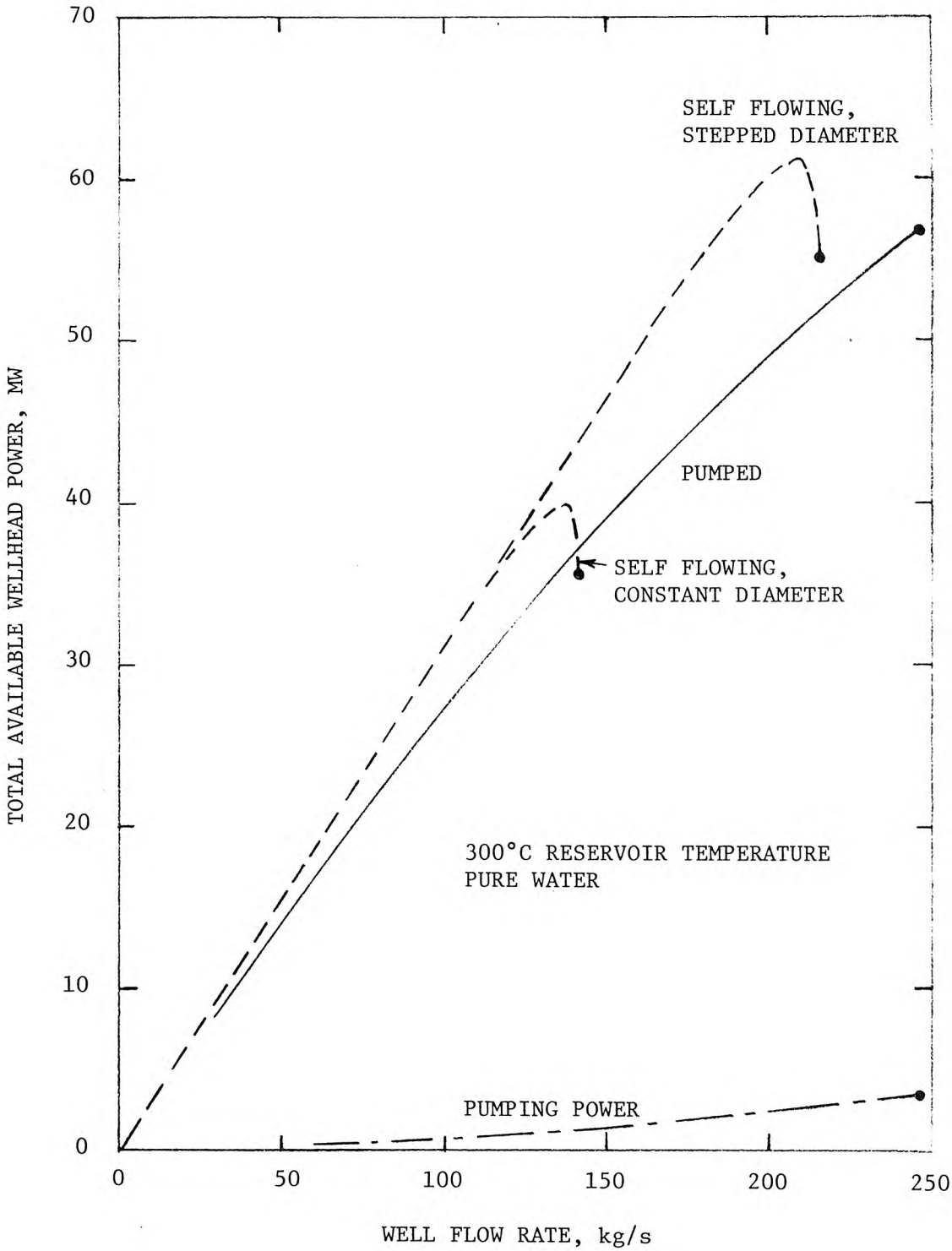

Fig. 7 Comparison of total available wellhead power with self flowing and pumping for 300°C reservoir temperature and pure water

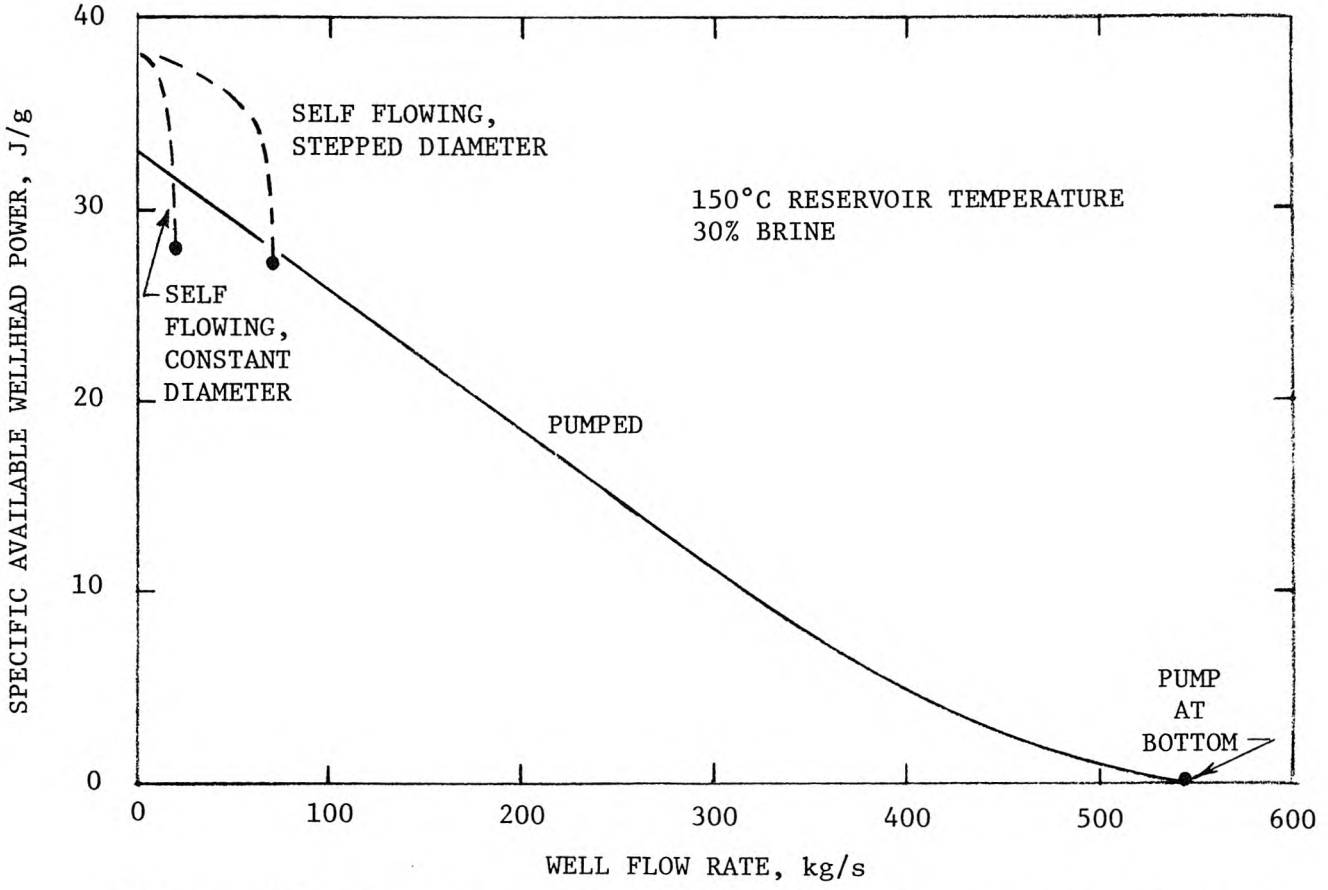

Fig. 8 Comparison of specific available wellhead power with self flowing and pumping for 150°C reservoir temperature and 30% brine

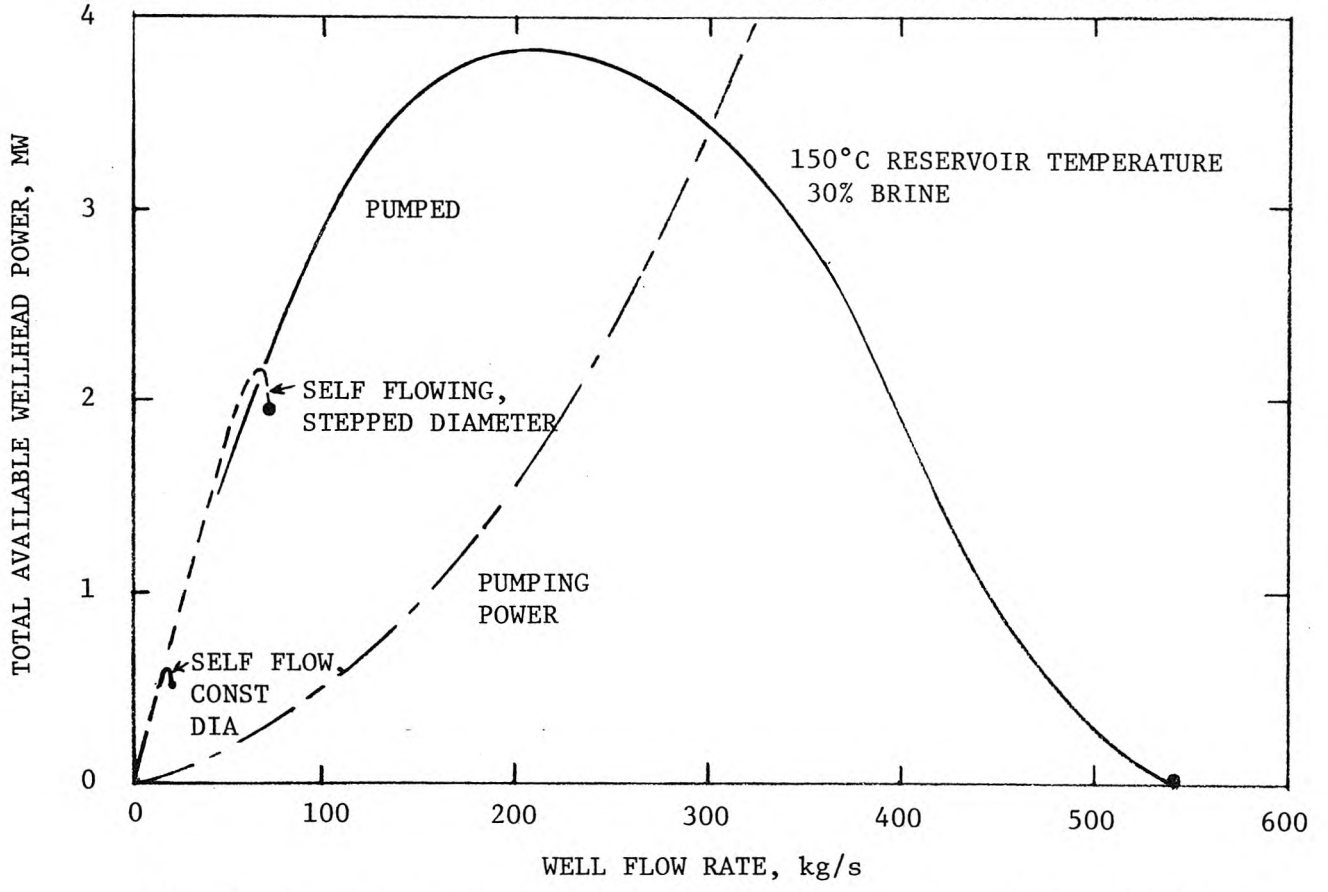

Fig. 9 Comparison of total available wellhead power with self flowing and pumping for 150°C reservoir temperature and 30% brine

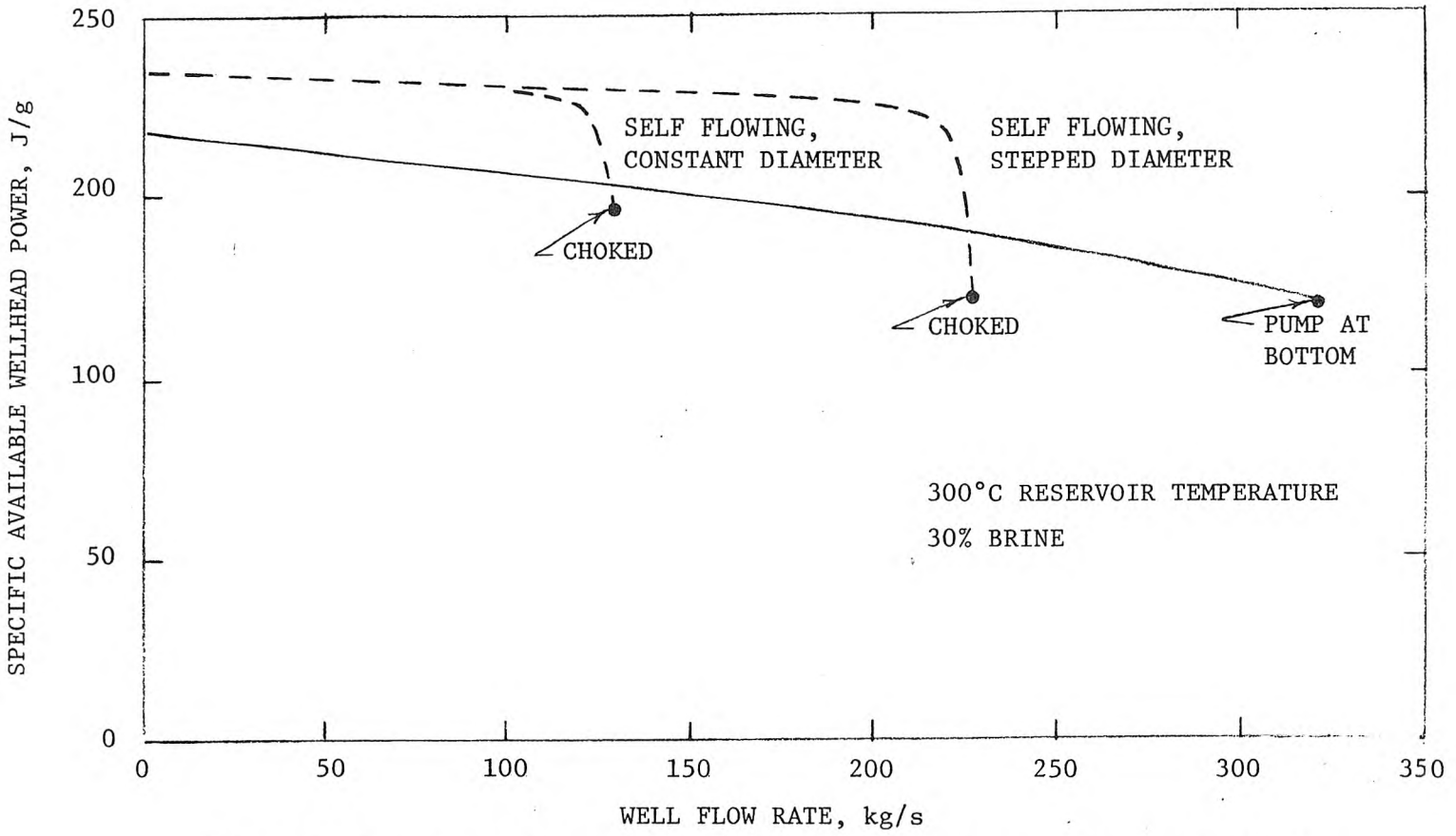

Fig. 10 Comparison of specific available wellhead power with self flowing and pumping for 300°C reservoir temperature and 30% brine

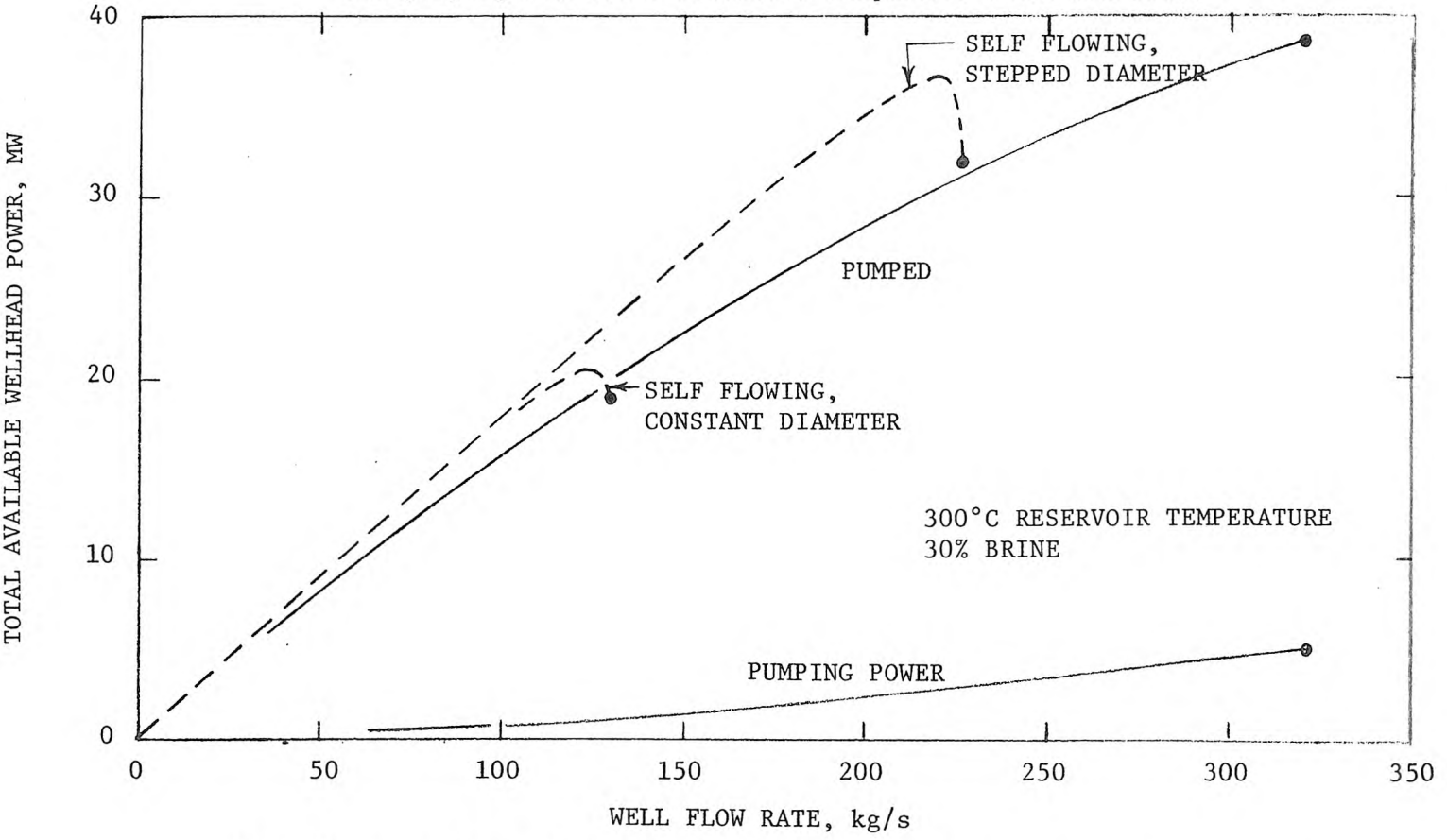

Fig. 11 Comparison of total available wellhead power with self flowing and pumping for 300°C reservoir temperature and 30% brine

Perhaps the largest pump that can be installed in a 0.25 m diameter casing would provide 1.0 MW pumping power, and a more realistic size would be 0.5 MW (the test pump described in Reference 1 is designed for 0.1 MW pumping power in a 0.22 m diameter casing). The flow rates at pumping powers (volume flow rate times pressure rise) of 0.5 and 1.0 MW are, therefore, adopted as representing the flow rates for near-term and advanced pumping systems, respectively.

Figure 12 compares the total wellhead available power for self flowing at peak-power flow rate in constant diameter and stepped diameter wells (dashed curves) with the total available wellhead power for pumping with 0.5 and 1.0 MW pumping power (solid curves) as a function of reservoir temperature for zero brine concentration. With self flowing the total available wellhead power increases by a factor of ten from 150°C reservoir temperature to 300°C reservoir temperature, and the power with stepped diameter is twice that with constant diameter. With pumping the total available wellhead power rises only half as much as with self flowing, and the available power with a 1.0 MW pump is only about 40 percent more than with a 0.5 MW pump. Compared with self flowing at constant diameter, the 1.0 MW pump provides higher powers at reservoir temperatures below 270°C and the 0.5 MW pump provides higher powers at reservoir temperatures below 240°C. Pumping provides little or no gain over self flowing with stepped diameter at any temperature.

Figure 13 presents the same comparisons with 30 percent brine concentration. The main difference is that with self flowing the total available power increases by a factor of twenty, instead of ten, from 150°C reservoir temperature to 300°C reservoir temperature, whereas the factor of increase with pumping is about the same as with pure water. The result is that at low temperatures pumping provides a greater increase in total available wellhead power, over self flowing, with brine than with pure water. However, no low-temperature, high-salinity geothermal resources are known.

The results for pumping, although calculated for a turbine-driven pump using cooling of the well flow, are also valid for a pump driven by mechanical, hydraulic, or electric power from the surface. In those cases, the power to drive the pump is

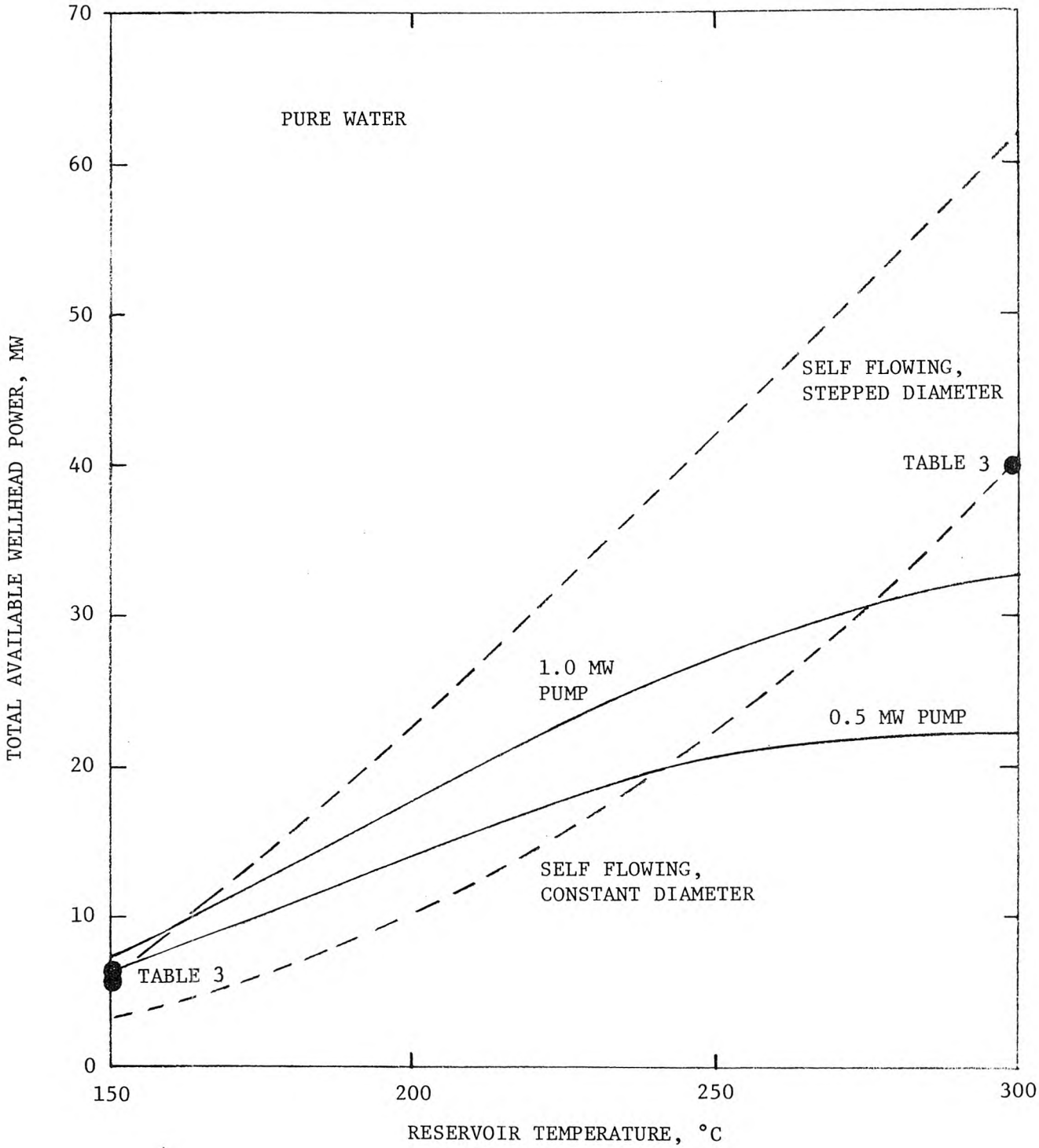

Fig. 12 Comparison of total available wellhead power for self flowing (at peaks) and pumping (at fixed pumping powers) as a function of reservoir temperature for pure water

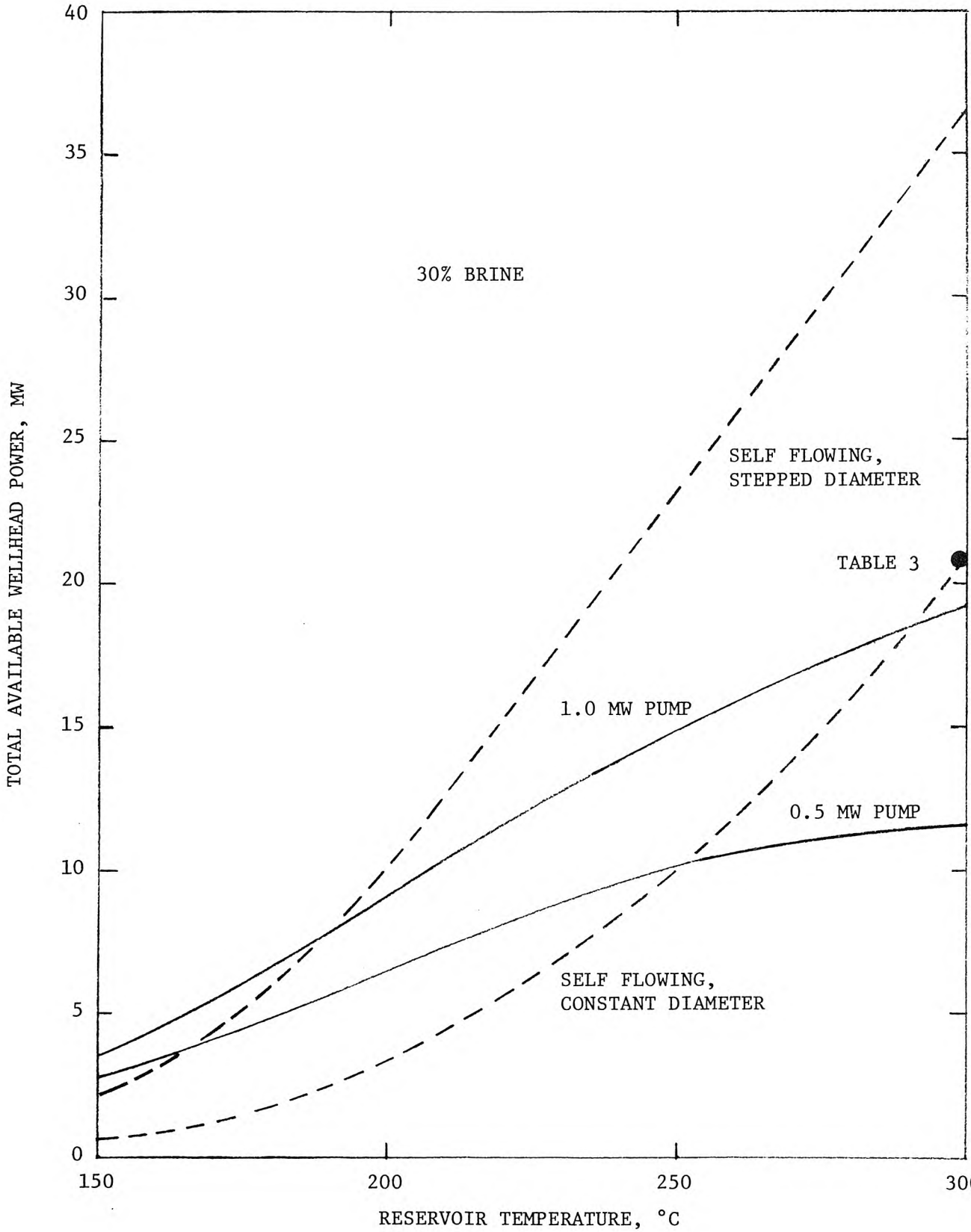

Fig. 13 Comparison of total available wellhead power for self flowing (at peaks) and pumping (at fixed pumping powers) as a function of reservoir temperature for 30% brine

taken from the conversion process output, and the resulting reduction in net output is about the same as the reduction in available wellhead power calculated here, for equal efficiency of conversion of heat to pumping power. For example, at 200°C reservoir temperature the flow rate with 1.0 MW pumping power is 175 kg/s, the pump pressure rise is 5.0 MPa (700 psi), and the wellhead temperature is 180°C (due to the assumed 4°C per MPa cooling). The wellhead available power is 17 MW. If the pump is externally driven, the wellhead temperature is 200°C and the wellhead available power is increased to 22 MW. The pump pressure rise is increased 10 percent by the higher saturation pressure, requiring 1.1 MW pumping power. If the pump efficiency is 0.5 and the available wellhead power is converted to pump-drive power at 0.5 efficiency, the amount of available wellhead power used for pumping is 1.1/(0.5 x 0.5) = 4.4 MW. This leaves 17.6 MW of available wellhead power, about the same as with the pump driven by cooling of the well flow.

The comparisons between self flowing and pumping are not significantly affected if the drawdown pressure or static well-bottom pressure is changed. Figure 14 shows how the curves in Figure 12 change if the drawdown pressure factor is increased by a factor of four. The wellhead powers are cut in half, but the relationship between self flowing and pumping is nearly the same.

The flow rates in Figure 14 range from 60 kg/s to 110 kg/s. Therefore Figure 14 also applies to a well with static bottom-hole pressure reduced to between 7.5 and 10.0 MPa, values that are only 51 to 68 percent of the hydrostatic pressure of ground water assumed in the previous calculations.

F. Summary

The comparisons between self flowing and pumping cast considerable doubt on the value of pumping. Above 225°C, pumping increases total available wellhead power by at most a factor of two compared with self flowing at constant diameter, and there is no increase compared with stepped diameter. At lower temperatures, pumping increases total available power by greater than a factor of two only with high brine concentrations that are not encountered at those temperatures.

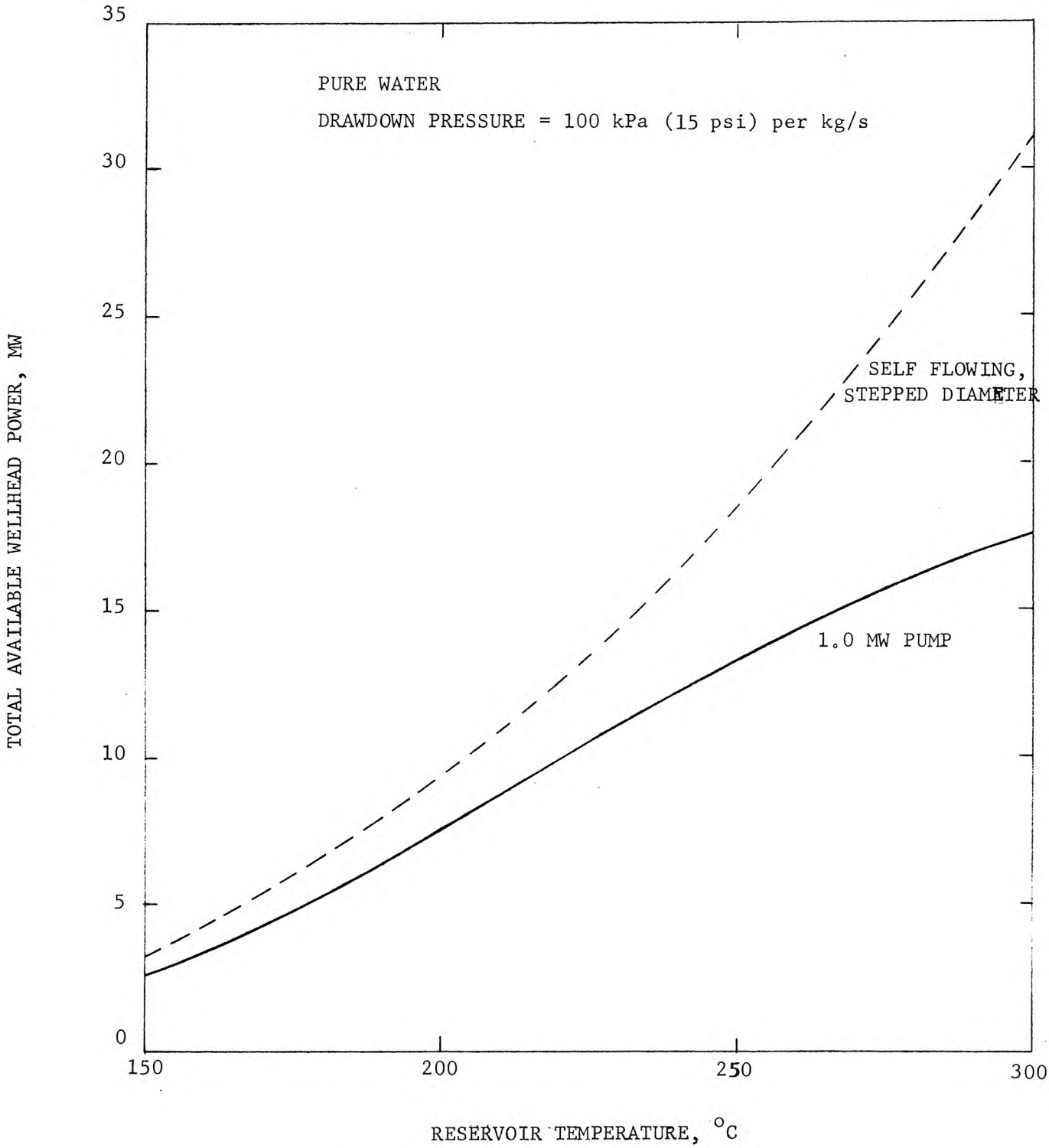

Fig. 14 Comparison of total available wellhead power for self flowing and pumping with four-fold increase in drawdown pressure

The possible value of pumping in preventing flashing and resulting scale deposit in the well remains, but it does not appear that pumping is of significant value on an energy basis.

III. CONVERSION PROCESSES

A. Wellhead Conditions

For the comparison of conversion processes, four wellhead conditions are chosen that are representative of the extremes calculated in the previous Section. The assumed wellhead conditions are given in Table 3 and also indicated on the curves of total available wellhead power in Figures 12 and 13.

The first wellhead condition is for self flowing at 150°C reservoir temperature with zero brine concentration and stepped diameter. The 150°C reservoir temperature represents the low end of the temperature range for geothermal electric power generation. Most brines at that temperature have salt concentration below 5 percent and can, therefore, be represented by pure water for power calculations. The stepped diameter requires enlarging the casing only to 300 m depth.

The second wellhead condition is for pumping at 150°C reservoir temperature and zero brine concentration with 0.5 MW pumping power. The pumping provides saturated liquid at the wellhead without vapor, and the 0.5 MW pumping power is a practical goal.

The third wellhead condition is for self flowing at 300°C reservoir temperature with zero brine concentration and constant diameter. The 300°C reservoir temperature represents the upper end of the temperature range available in geothermal reservoirs. Some brines at that temperature have low salt concentrations and can be represented by pure water for power calculations. A constant diameter provides high flow rate and stepped diameter is not needed.

The fourth wellhead condition is the same as the third except that 30 percent brine concentration is assumed.

Table 3 Wellhead Conditions for Conversion Process Calculations

	150°C reservoir temperature, self flowing, stepped diameter, pure water	150°C reservoir temperature, pumped, 0.5 MW pumping power, pure water	300°C reservoir temperature, self flowing, constant diameter, pure water	300°C reservoir temperature, self flowing, constant diameter, 30% brine
Temperature, °C	114	137	218	202
Quality	0.07	0	0.21	0.12
Pressure, MPa	0.16	0.33	2.23	1.25
Pressure, psi	23	48	323	181
Flow rate, kg/s	116	137	136	124
Specific available power, J/g	54	48	292	171
Total available power, MW	6.3	6.6	40	21

B. Flash Steam Process

The most commonly used conversion process for power generation from brine is the flash steam process. Figure 15 shows a flash steam process with one stage. The flow leaving the wellhead enters a flash vaporizer where the pressure is reduced, causing part of the brine to vaporize. The mixture of brine and steam then enters a separator where the brine and steam are separated. The brine flows to a reinjection well or other disposal area. The steam flows to a turbine for power generation and is then condensed in a condenser. The condensate is pumped to atmospheric pressure for disposal or for use in the cooling tower.

The flow from a pumped well consists entirely of liquid (designated L in the diagram), and the flow from a self flowing well consists of both liquid (L) and gas (G), the gas phase being steam. With two-phase wellhead flow the flash vaporizer may be omitted and the wellhead flow fed directly to the separator at wellhead pressure. The flash vaporizer, if used, consists of whatever provides the pressure drop between the wellhead and the separator; usually an orifice at the separator inlet is used.

The separator is any conventional type such as a cyclone separator. With high brine concentrations the steam line to the turbine may include scrubbers to remove entrained salts. The condenser is usually a contact condenser wherein water from the cooling tower is sprayed directly into the condensing steam. The condenser pumping may be accomplished by elevating the condenser above ground level or by using a mechanical pump. Any noncondensables in the steam must be pumped from the condenser to atmosphere by mechanical pumps or by steam ejectors using steam that would otherwise drive the turbine.

A flash steam process with two stages is shown in Figure 16. The brine from the first-stage separator flows to the second-stage flash vaporizer where the pressure of the brine is reduced to that of the steam leaving the first-stage steam turbine. The resulting steam and brine mixture is separated in the second-stage separator. The steam from the second-stage separator joins the steam leaving the first-stage turbine and flows to the second-stage turbine.

To minimize moisture in the second-stage turbine, the moisture leaving the first-stage turbine is separated and added to the brine leaving the second-stage

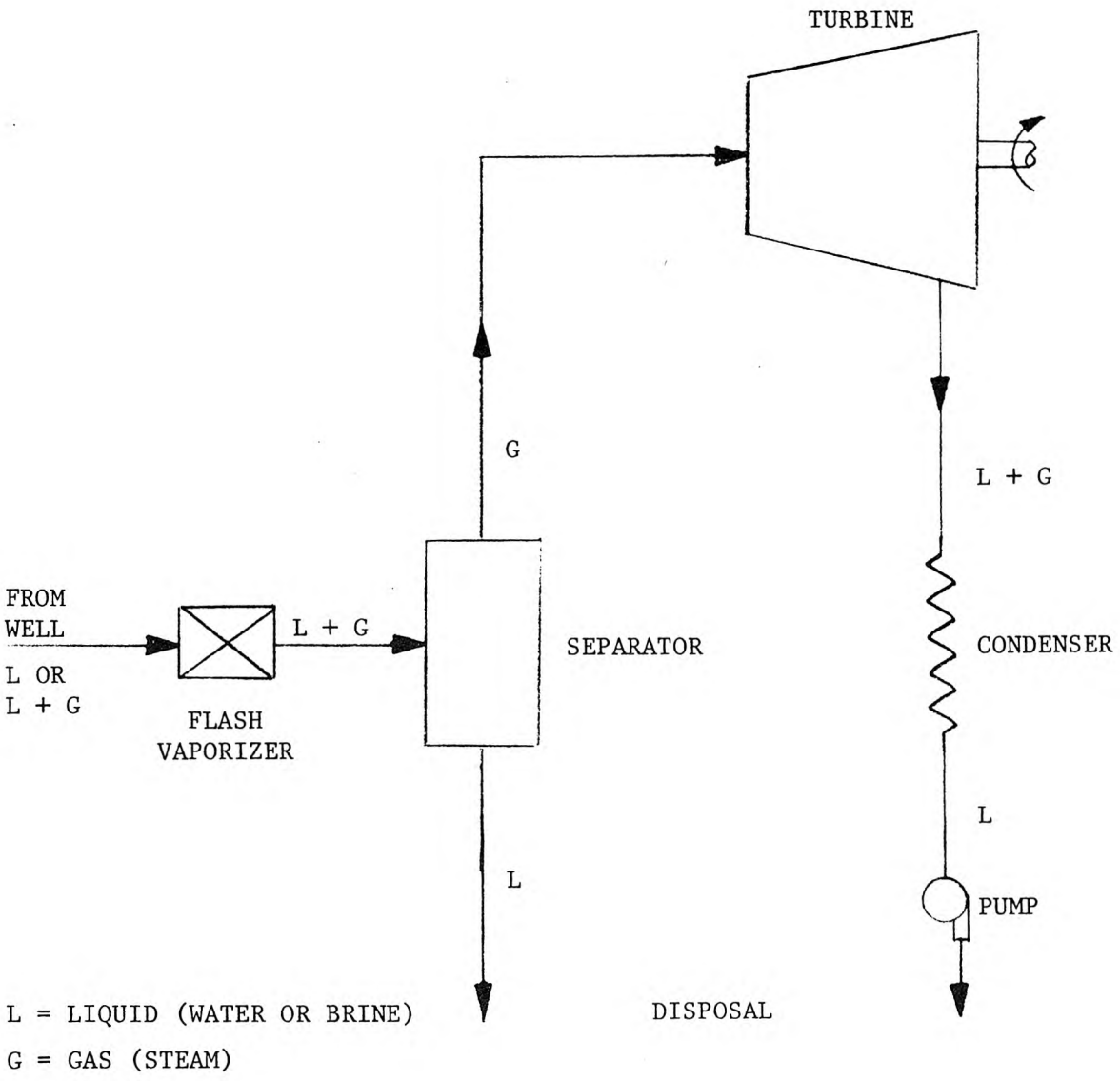

Fig. 15 Flash steam process with one stage

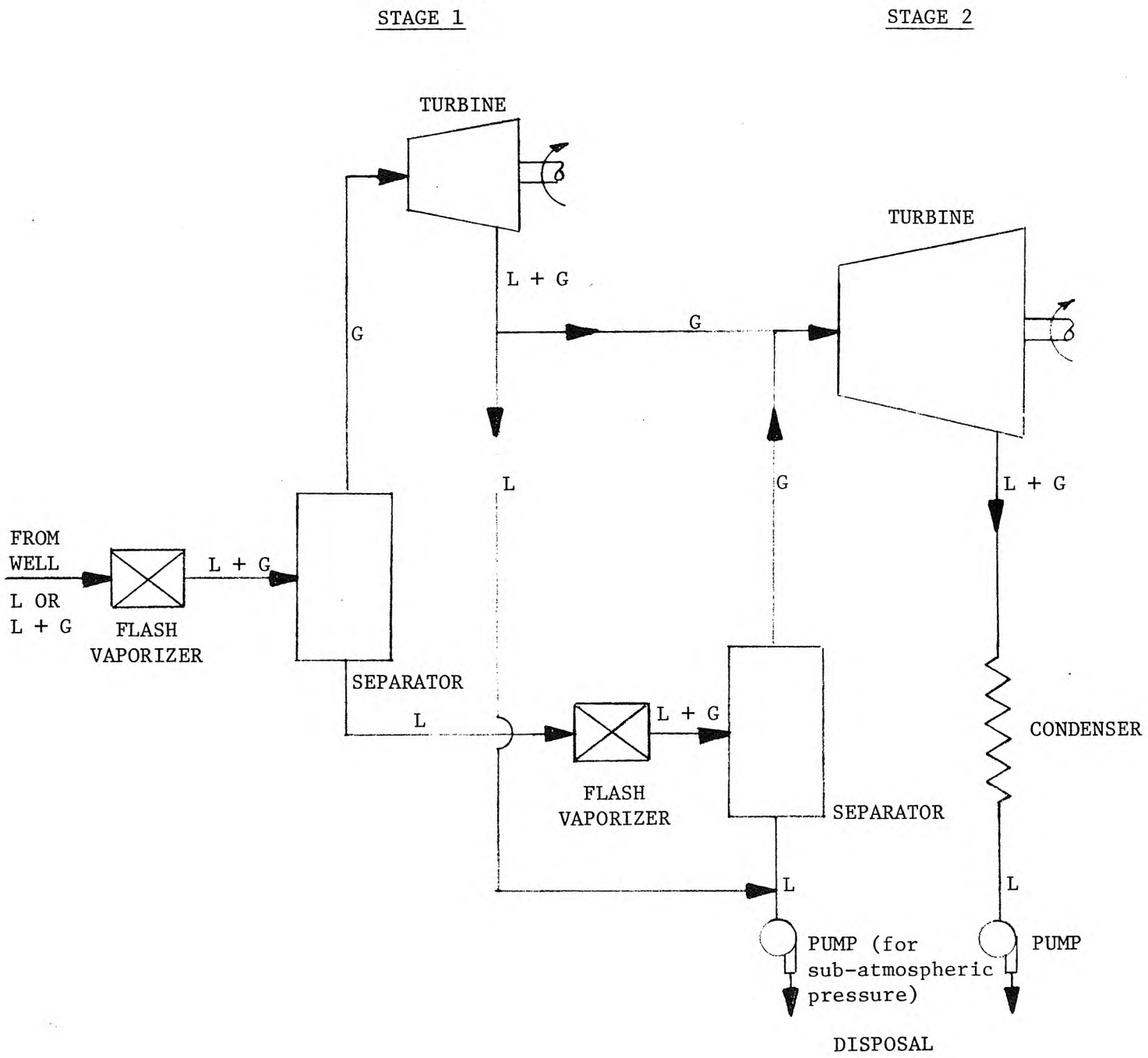

Fig. 16 Flash steam process with two stages

separator. If the pressure in the second-stage separator is below atmospheric, a pump (or elevation above ground level) is required for removal of the separated brine.

A flash steam process with three stages is shown in Figure 17. The pattern is the same as for the two-stage process. The gain achieved with multi-staging is that the brine leaving the last-stage separator is rejected at lower temperatures as the number of stages is increased, resulting in more of the brine energy being converted to turbine output power.

C. Dual Steam Process

When brine, or a brine and steam mixture, is flash vaporized, mechanical work is available from the expansion of the vaporizing mixture. That work is not utilized in the flash steam system; the work goes into accelerating the flow and is dissipated as heat when the flow is decelerated by friction and impact. It is possible to replace the flash vaporizer with a two-phase expander that provides the same pressure drop but extracts part of the available work as shaft power.

Figure 18 shows a single-stage process using a two-phase expander. The process is the same as the one in Figure 15 except that a two-phase expander replaces the flash vaporizer. The two-phase expander provides shaft power that can be used to generate additional electric power. At the same separator pressure as in the flash steam process, the exhaust from the two-phase expander has lower quality than the exhaust from the flash vaporizer, and the steam turbine power is, therefore, reduced. But the sum of the steam turbine power and two-phase expander power is always greater than for the flash steam process no matter how low the two-phase expander efficiency. The flash vaporizer can be considered as the limiting case of a two-phase expander with zero efficiency.

Since two types of expanders are being used in combination, the process of Figure 18 is designated as a dual steam process. The dual steam process was first proposed by Laird.[5]

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[5] Laird, A. D. K., "Water from Geothermal Resources," Geothermal Energy, (Kruger, P., and Otte, C., ed.), pp. 178-196, Stanford University Press, Stanford, California, 1973.

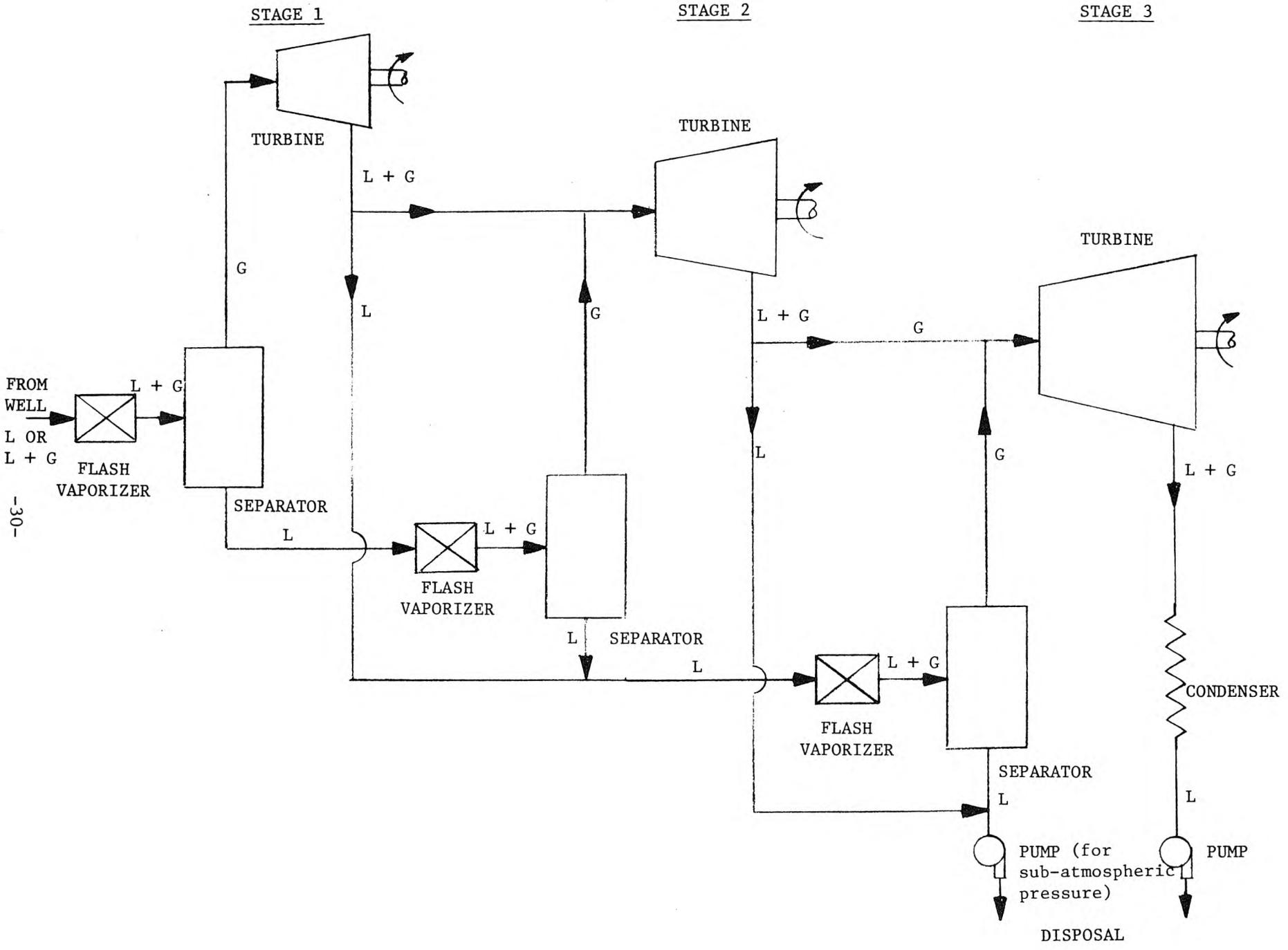

Fig. 17 Flash steam process with three stages

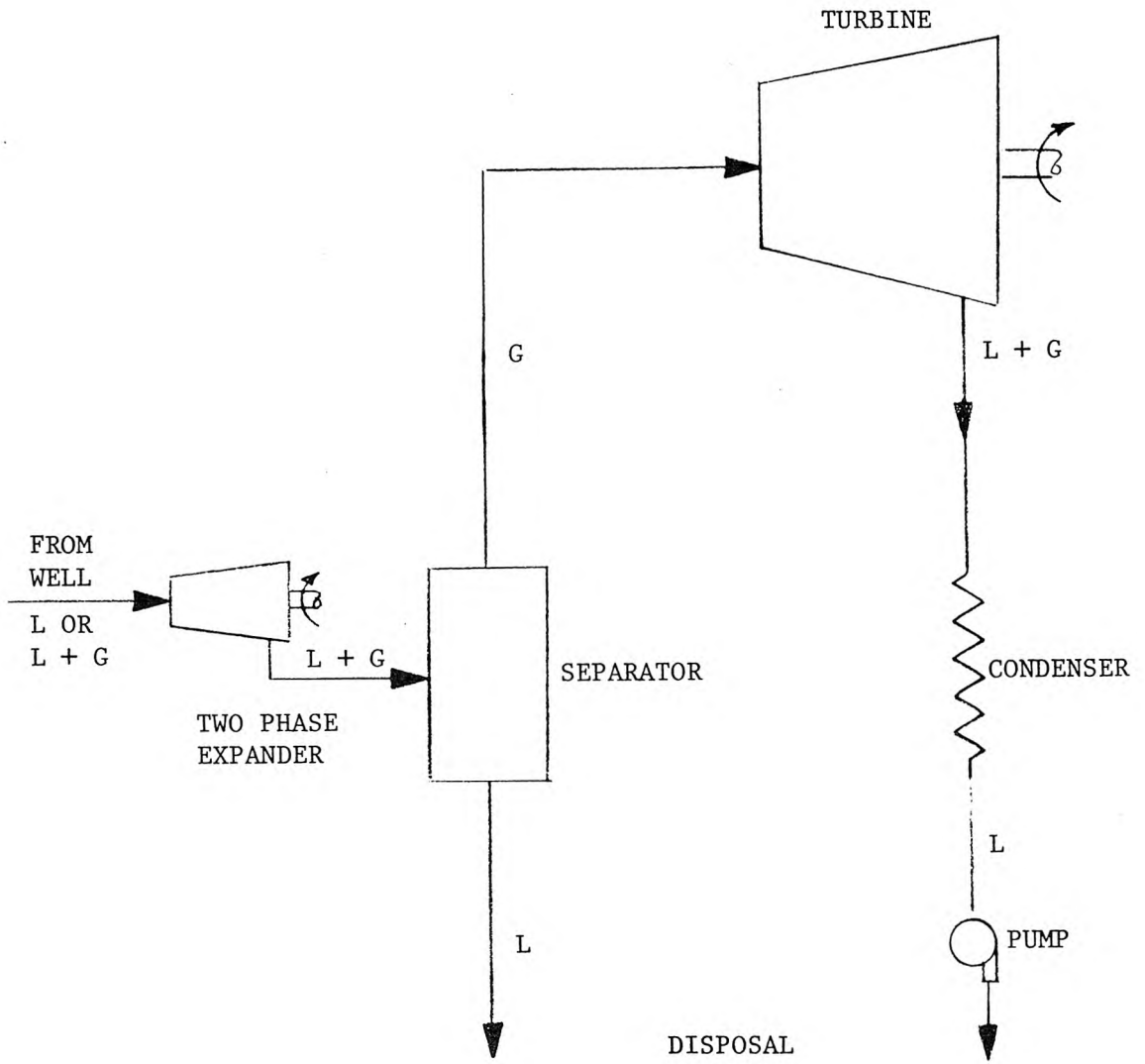

Fig. 18 Dual steam process with one stage

A dual steam process with two stages is shown in Figure 19. As in the case of the flash steam process, the gain achieved with multistaging is that the brine leaving the last-stage separator is rejected at lower temperatures as the number of stages is increased.

The liquid leaving the last-stage separator is still at a higher temperature and pressure than the condenser, and additional mechanical work is available from expanding that liquid through an additional two-phase expander to the condenser conditions. A single-stage dual steam process with an exhaust liquid expander is shown in Figure 20. The added expander is shown exhausting into a separator at the condenser pressure, with the exhaust steam flowing to the condenser inlet and the exhaust liquid flowing to the condenser exit. This avoids problems of brine contamination of the condenser and cooling system.

With the exhaust liquid expander, all of the wellhead flow is rejected at condensing temperature. The process extracts all of the available energy from the wellhead flow, within the limitations of mechanical efficiencies, using only one steam turbine and two two-phase expanders. However, from the standpoint of limiting the volume flow rate in the two-phase expanders more stages may be desired, as discussed later. A two-stage dual steam process with an exhaust liquid expander is shown in Figure 21.

D. Total Flow Process

Another method of expanding the entire wellhead flow to condensing temperature and extracting the available work is the total flow process.[6] In the total flow process, shown in Figure 22, the entire wellhead flow expands through one or more two-phase expanders to the condenser conditions. Multistaging of the two-phase expanders may be used for mechanical reasons, but thermodynamically all of the recoverable energy can be obtained in a single stage.

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[6] Austin, A. L., Higgins, G. H., and Howard, J. H., The Total Flow Concept for Recovery of Energy from Geothermal Hot Brine Deposits, Report No. UCRL-51366, Lawrence Livermore Laboratory, Livermore, California, April 3, 1973.

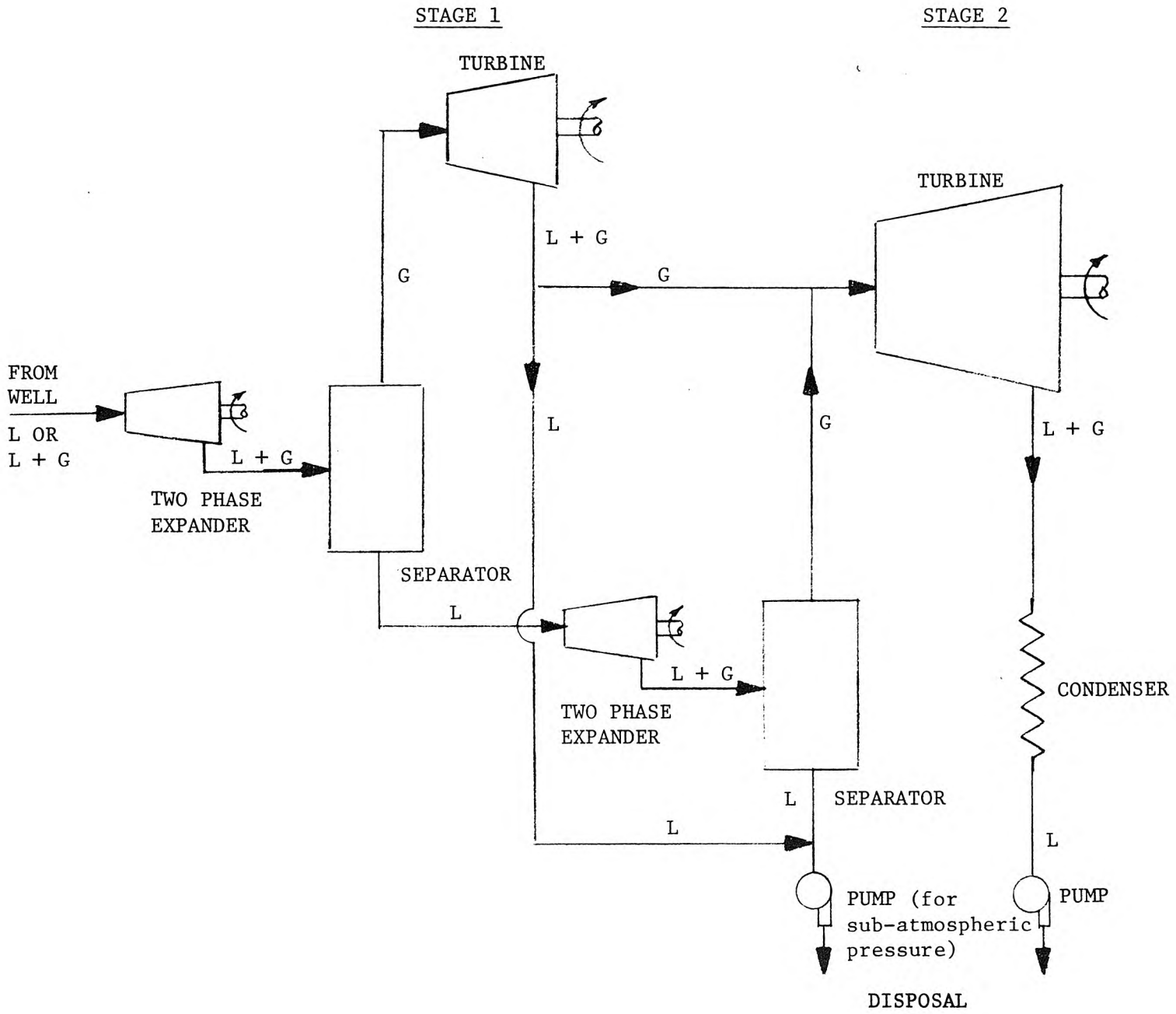

Fig. 19 Dual steam process with two stages

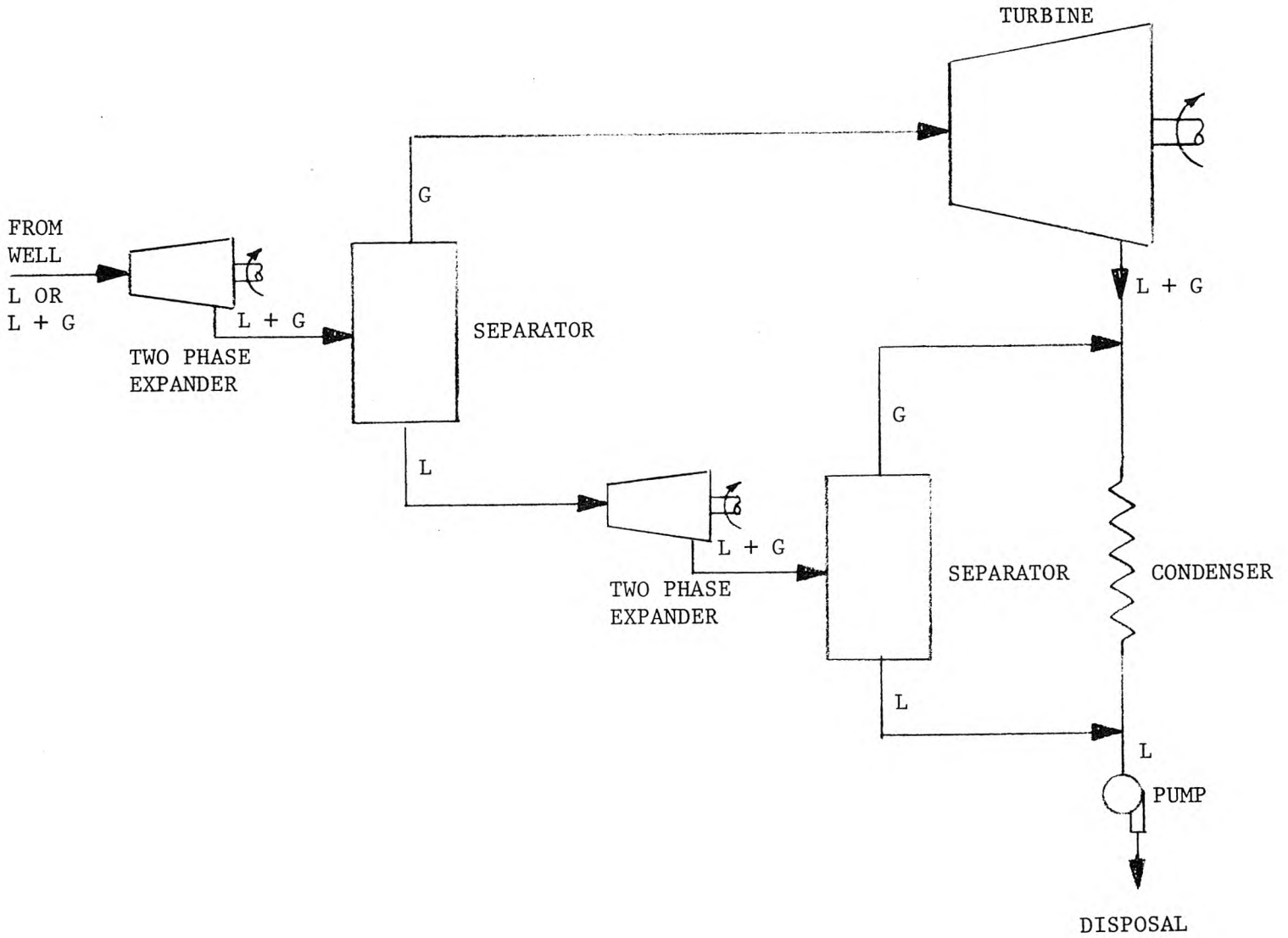

Fig. 20 One-stage dual steam process with exhaust liquid expander

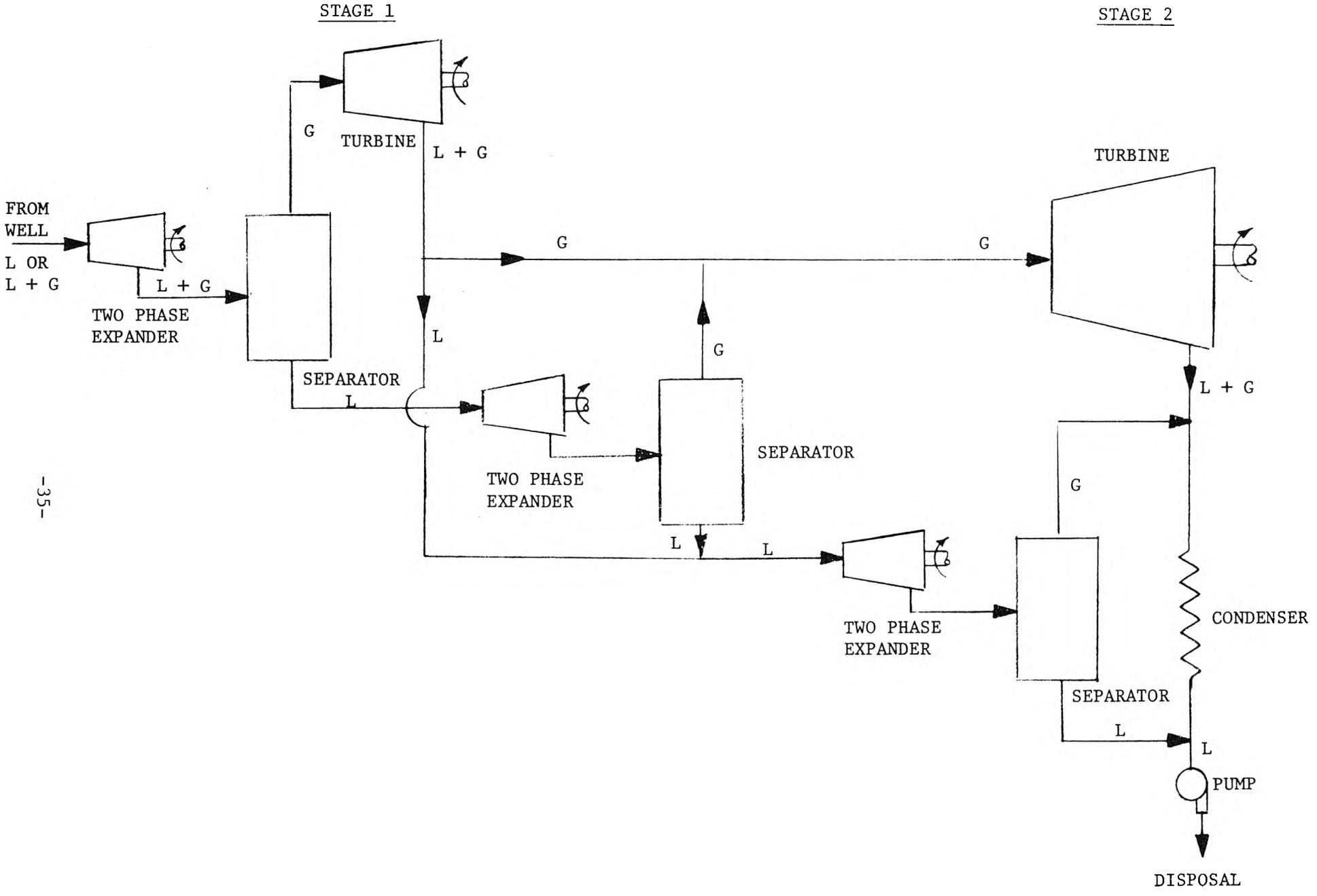

Fig. 21 Two-stage dual steam process with exhaust liquid expander

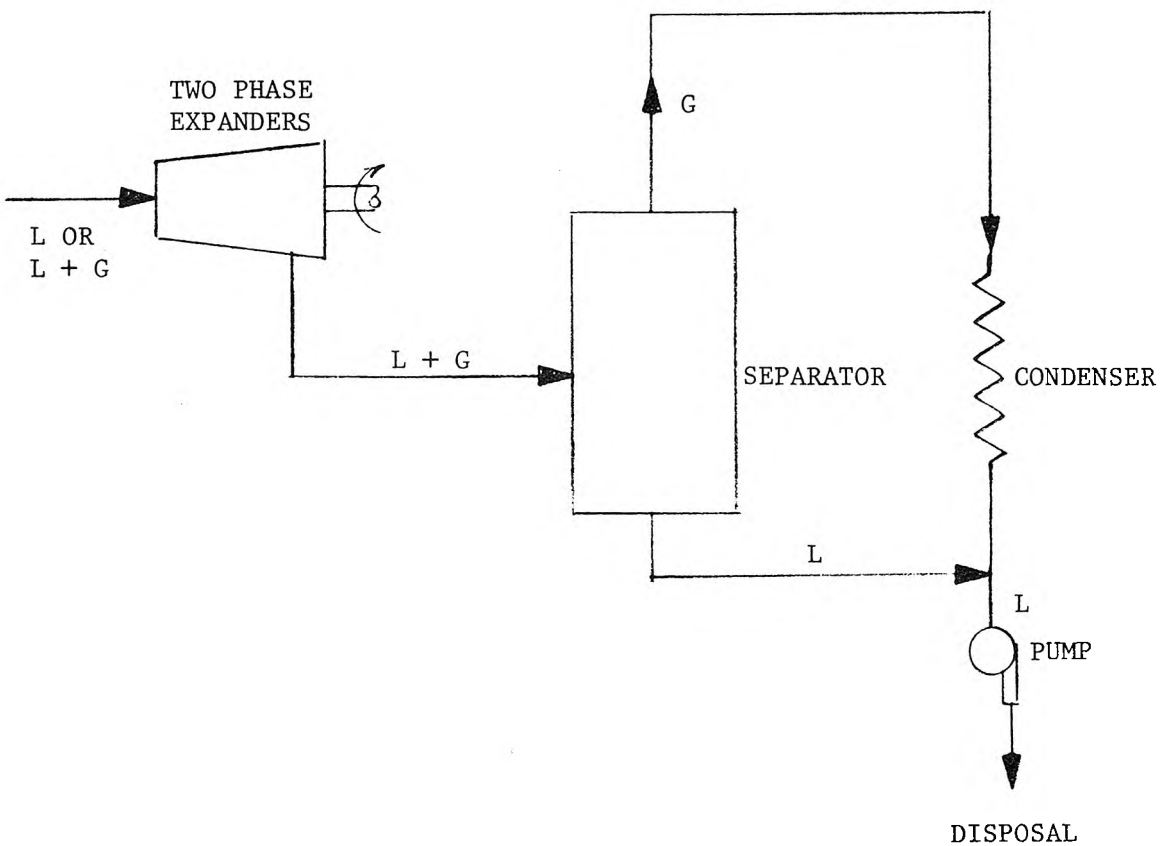

Fig. 22 Total flow process

E. Binary Process

The binary conversion process uses the wellhead flow as a heat source for a separate closed-loop power cycle. Usually an organic working fluid operating in a Rankine cycle is considered. Figure 23 shows a binary process. The wellhead flow, if it is two-phase, enters a separator, and the steam leaving the separator enters the steam heat exchanger where the steam is condensed by heat trnasfer to the organic working fluid. The condensate then joins the brine leaving the separator and flows through the liquid heat exchanger where the liquid is cooled by further heat exchange to the organic working fluid. The liquid then flows to disposal, through a pump if necessary. If the wellhead flow is single phase, as with a pumped well, then no separator or steam heat exchanger is used.

The organic working fluid flows from the cold end to the hot end of the liquid heat exchanger and then through the steam heat exchanger, if used. The organic fluid then expands through the turbine, is condensed by heat transfer to atmosphere or to a cooling fluid, and is pumped back to the heat exchanger inlet.

A single-stage binary process such as shown in Figure 22 can cool all of the brine to condensing temperature plus some minimum heat exchanger temperature difference, and multistaging is not needed.

F. Flash Binary Process

High salt concentrations may cause fouling of the liquid heat exchanger in the binary process. The heat in the brine can still be recovered, however, by flash vaporizing the brine and using the steam in a series of steam heat exchangers. Such a flash binary process with one stage is shown in Figure 24. The wellhead flow, after possible additional vaporization in a flash vaporizer, is separated, and the steam is used to heat the organic working fluid. The separated brine flows to the reinjection well or other disposal site.

The heat exchanger in Figure 24 is both a steam and liquid heat exchanger. After the steam condenses in the first part of the heat exchanger the condensate

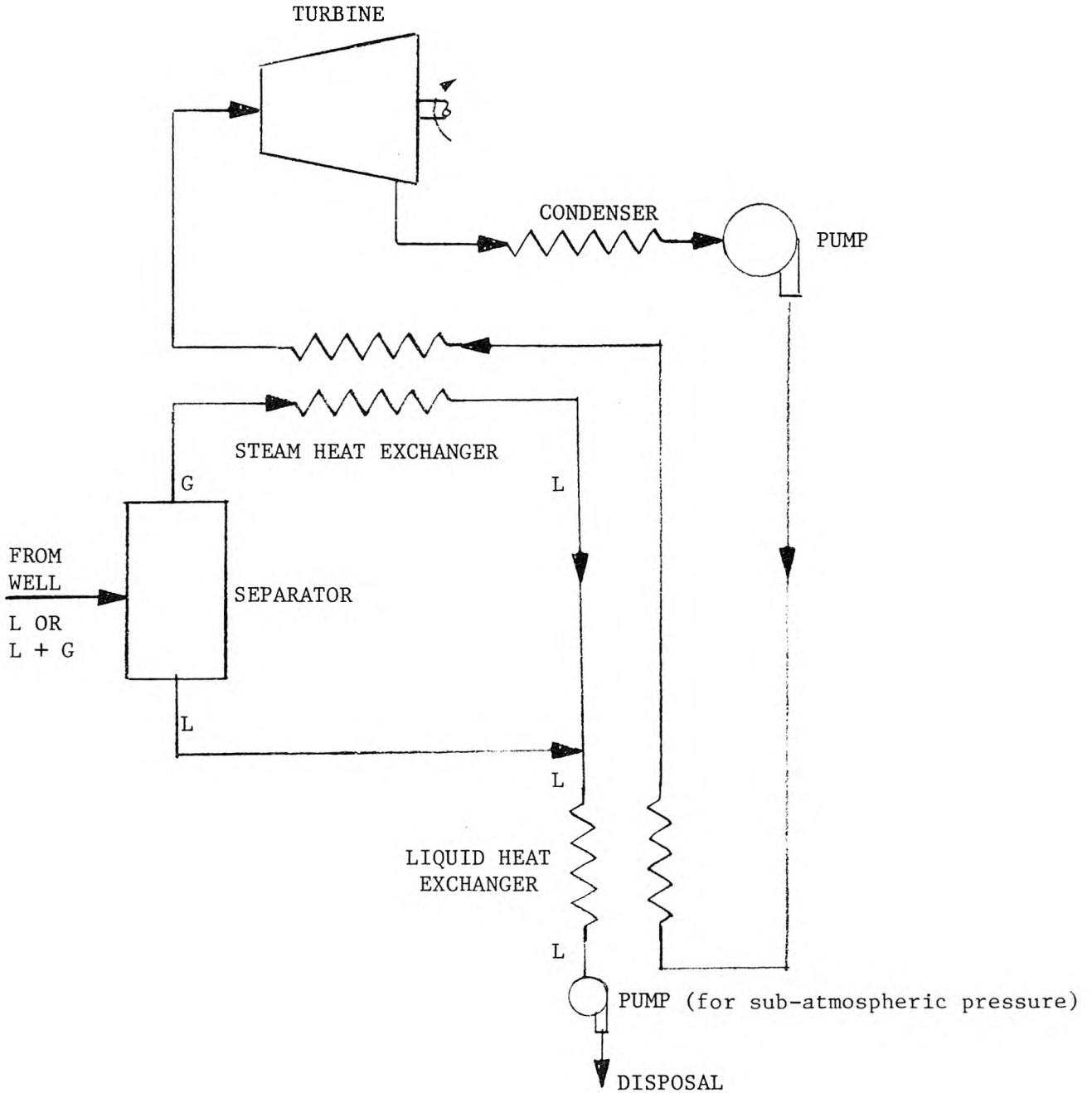

Fig. 23 Binary process

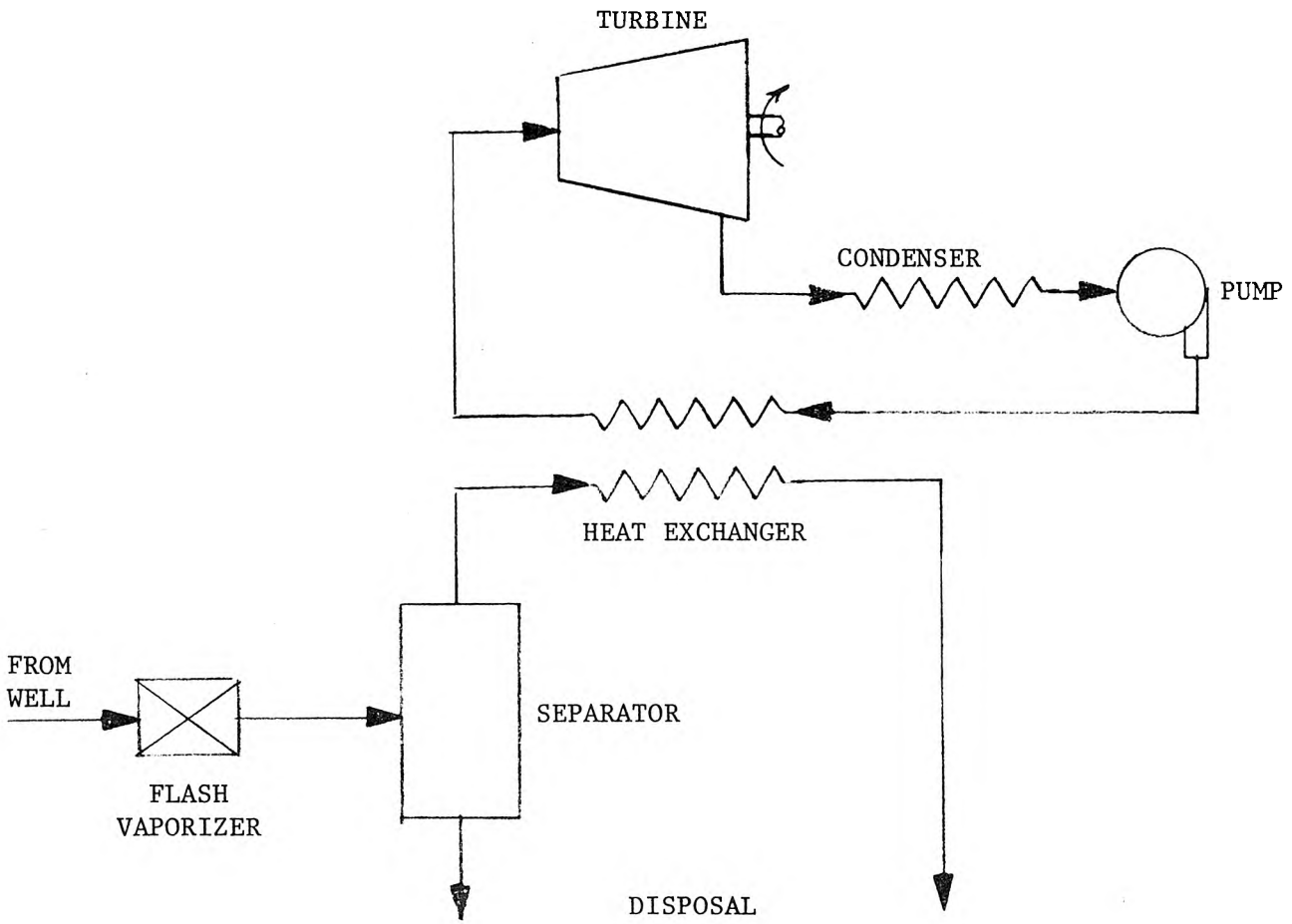

Fig. 24 Flash binary process with one stage

(which is pure water) is cooled by further heat transfer to the organic working fluid.

A flash binary process with two stages is shown in Figure 25. The steam entering the first-stage heat exchanger is condensed, and the condensate is cooled to an intermediate temperature. The condensate is then throttled to saturation pressure (the mechanical work lost is negligible). The brine from the first-stage separator flows to the second-stage flash vaporizer where the pressure of the brine is reduced to that of the liquid leaving the first-stage heat exchanger. The resulting steam and brine mixture is separated in the second-stage separator. The steam from the second-stage separator joins the liquid leaving the first-stage heat exchanger and flows to the second-stage heat exchanger. The brine from the second-stage separator flows to disposal, through a pump if necessary. The liquid leaving the second-stage heat exchanger flows to disposal or is used in the cooling tower.

G. Method of Calculation

The method of calculation for flash steam and dual steam is derived in Appendix D. The quantities specified are wellhead temperature, quality, and brine concentration; number of stages; steam condensing temperature; steam turbine and two-phase expander efficiencies; disposal and condenser pump efficiencies; and selection of optional exhaust liquid expander. The calculation procedure first assigns initial guesses to the turbine inlet temperatures in each stage. The flash vaporizer, or two-phase expander, exit conditions are then calculated stage by stage together with the steam turbine exit conditions, steam turbine power output, and any two-phase expander power output. The powers for all of the stages are added to give the total electric power output. The calculations are then repeated with other turbine inlet temperatures until the optimum temperatures corresponding to maximum total output power are found, using a standard multivariate search routine.

The method of calculation for total flow is simply to multiply the wellhead available power by the assumed two-phase expander efficiency to obtain the output power.

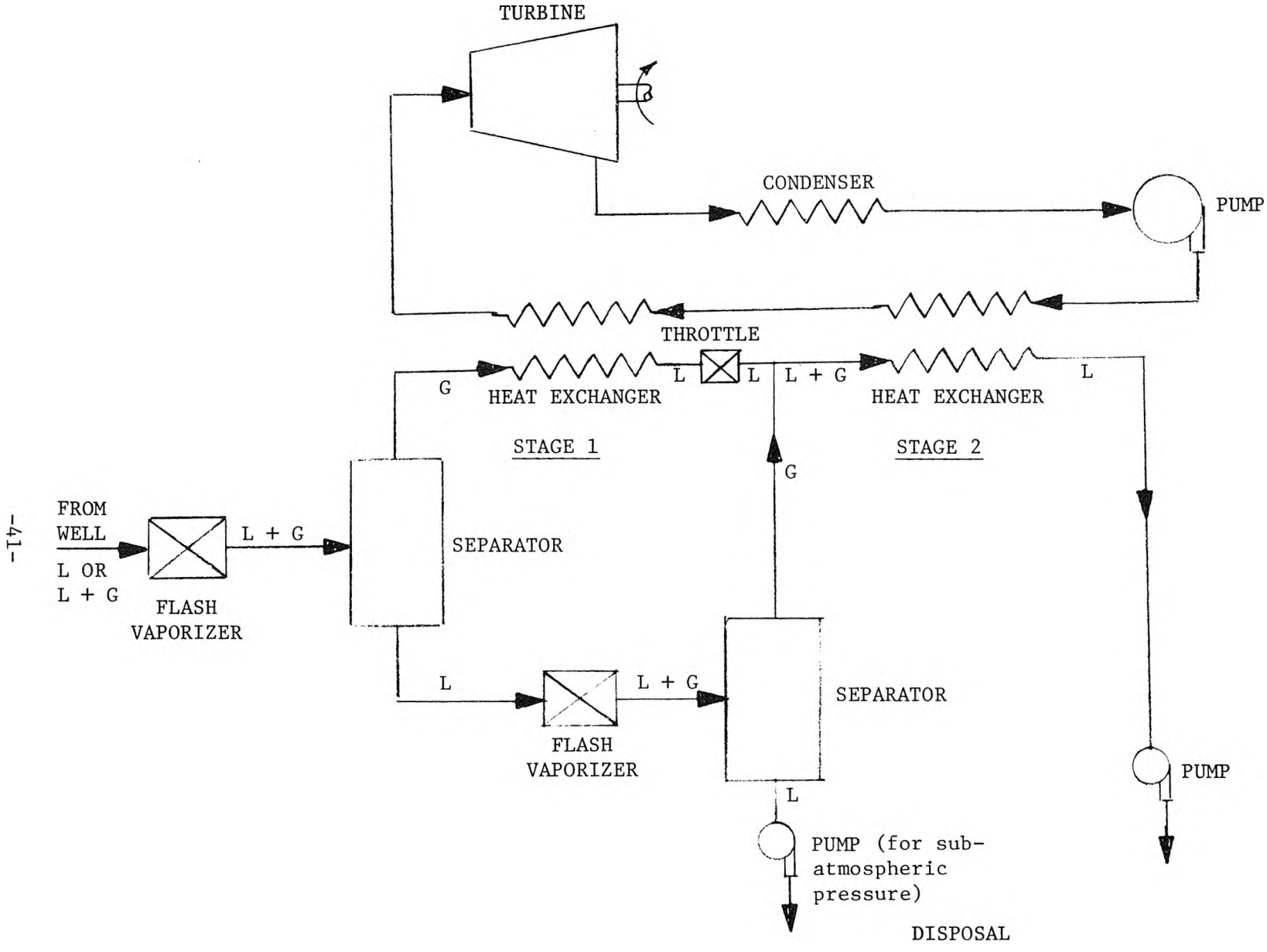

Fig. 25 Flash binary process with two stages

The method of calculation for binary and flash binary is derived in Appendix E. The quantities specified are wellhead temperature, quality, and brine concentration; number of stages; organic working fluid condensing temperature; disposal pump efficiency; temperature difference between the water or steam and the organic fluid at the inlets and exits of the heat exchangers; and selection of a liquid heat exchanger to give a binary, instead of flash binary, process.

The losses in the organic cycle are lumped together in a cycle efficiency factor which is the ratio of actual to ideal power output, the ideal power output being the output of a series of Carnot cycles receiving heat along the temperature curve of the organic fluid as it proceeds through the heat exchangers and rejecting heat at the condensing temperature of the organic fluid. The cycle efficiency factor is chosen to reflect turbine and pump losses and the effect of any heat rejection above the condensing temperature.

The calculation procedure for binary and flash binary processes first assigns an initial guess to the inlet temperature of the first-stage heat exchanger. The temperatures at the inlets of the other heat exchangers (if more than one stage is used) are then calculated, together with the organic working fluid flow rate, to set the temperature differences between the organic working fluid and brine at the inlets and exits of the heat exchangers to the specified temperature difference. The ideal power output of the organic cycle is then calculated. The inlet temperature of the first-stage heat exchanger is varied and the calculations repeated until the optimum temperature corresponding to maximum power output is found.

H. Conversion Process Constants

Table 4 presents the constants used in the conversion process calculations. For the flash steam, dual steam, and total flow processes the steam condensing temperature is assumed to be 45°C, a typical value for steam power plants. The combined efficiency of the steam turbines and electric generators (the ratio of electric output to isentropically-available power) is assumed to be 0.7, a reasonable value with saturated vapor at the turbine inlets.

Table 4 Conversion Process Constants

A. Steam Processes

Steam condensing temperature	45°C
Steam turbine efficiency	0.7
Two-phase expander efficiency	0.6
Pump efficiency	0.7
Disposal pressure	0.1 MPa (1 atm)

B. Binary Processes

Organic fluid condensing temperature	45°C
Organic working fluid	isobutane
Pressure of organic working fluid in heat exchangers	4.14 MPa (600 psi)
Cycle efficiency factor	0.7
Temperature difference between water and organic working fluid at inlets and exits of heat exchangers	10°C (flash binary or binary with self-flowing well) 20°C (binary with pumped well)
Water disposal pump efficiency	0.7
Disposal pressure	0.1 MPa (1 atm)

The efficiency of two-phase expanders can be expected to be less than that of steam turbines, and a value of 0.6 is assumed. The efficiency of disposal and condenser pumps is assumed to be 0.7. Losses due to pressure drops are considered to be included in the turbine and two-phase expander efficiencies.

For the binary and flash binary processes, the organic fluid condensing temperature is assumed to be 45°C, the same as for the flash steam, dual steam, and total flow processes. This is a favorable assumption for the binary processes since the organic fluid must use a surface condenser whereas steam can use a contact condenser.

The only information needed on the organic fluid for the calculation method used here is the shape of the temperature-versus-enthalpy curve for calculating the heat exchanger temperatures on the organic side. For this purpose the enthalpy curve of isobutane at 4.14 MPa (600 psi) is used, which gives results representative of organic fluids at desirable operating conditions.

The value of the cycle efficiency factor is assumed to be 0.7, which corresponds to turbine and pump efficiencies of about 0.8 for an isobutane cycle, as shown in Appendix E, and is a reasonable value for any organic cycle.

The temperature difference between the water or steam and the organic fluid at the inlets and exits of the heat exchangers is specified to be 10°C for flash binary or binary with a self-flowing well. For binary with a pumped well a higher temperature difference is required, as will be shown, and 20°C is specified.

I. Comparison of Power Outputs

The results of the conversion process calculations are expressed as the power recovery fraction, which is the ratio of actual output power to available wellhead power.

Figure 26 compares the power recovery fraction of the different conversion processes for 150°C reservoir temperature and self flowing; the flow to the conversion processes is water and steam at 114°C and 0.07 quality (Table 3). The power recovery fractions are plotted as a function of the number of stages.

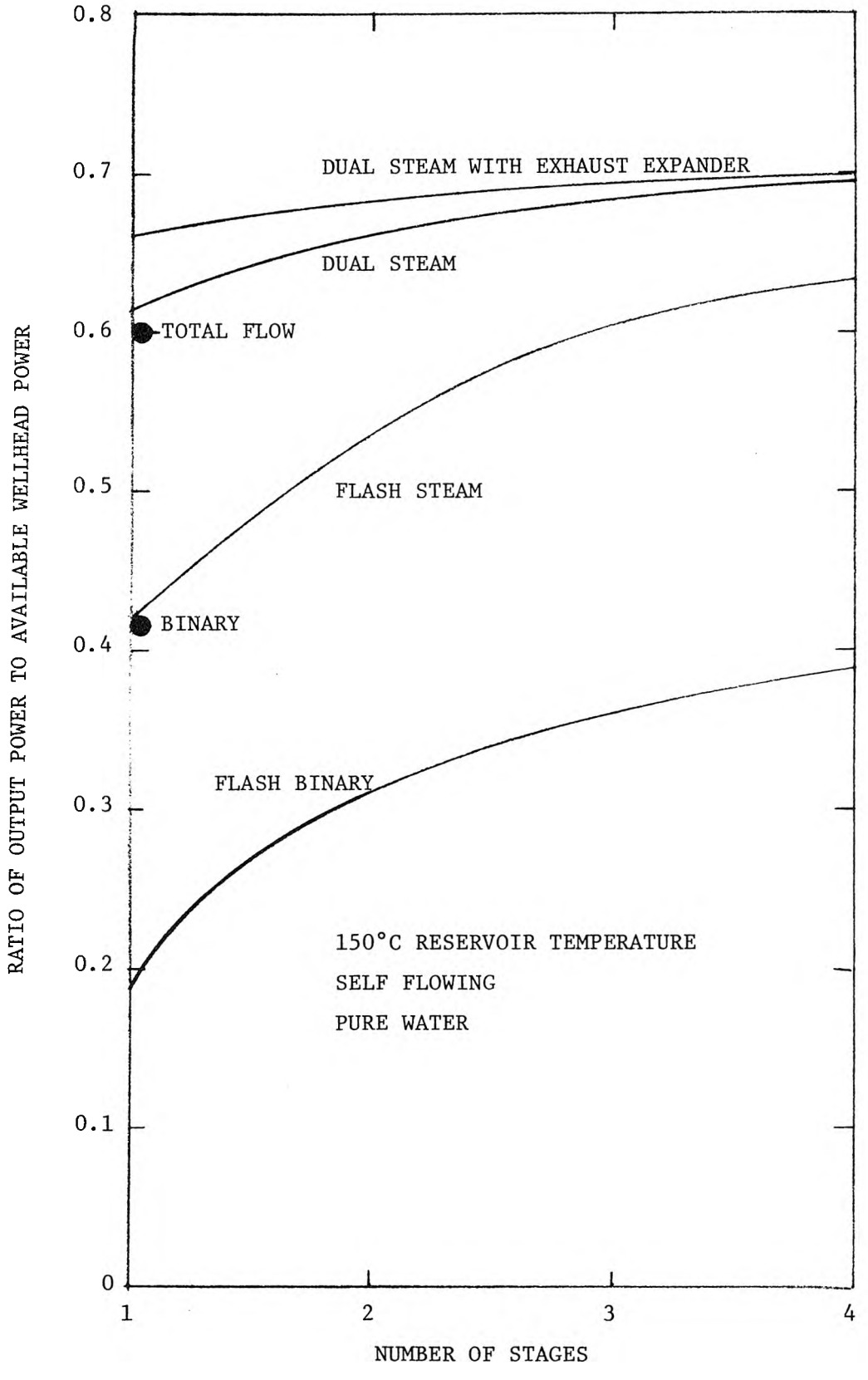

Fig. 26 Comparison of power recovery fractions for 150°C reservoir temperature, self flowing, pure water

The highest power recovery fraction is obtained with the dual steam process using an exhaust liquid expander. The power recovery fraction varies from 0.66 with one stage to 0.70 with four stages. The reason for the increase is that the portion of the output power produced by the more efficient steam turbines increases from 60 percent with one stage to 90 percent with four stages.

The power produced by the dual steam system without an exhaust liquid expander is only slightly less. The power recovery fraction varies from 0.62 with one stage to 0.69 with four stages.

The total flow process has a power recovery fraction of 0.6, equal to the assumed efficiency of the two-phase expanders. The total flow process gives slightly less power than the single-stage dual steam process because the latter generates 35 percent of its power in the assumed more-efficient steam turbine.

The power recovery fraction of the flash steam process is next highest, varying from 0.42 with one stage to 0.63 with four stages. Thus, the dual steam and total flow processes give 50 percent more power than the flash steam process, as claimed for total flow in Reference 6, but only when compared with a single-stage flash steam process. Compared with a flash steam process of three or more stages, the total flow process gives the same or less power (at the assumed 60 percent efficiency of two-phase expanders), and the dual steam process gives only about 10 percent more power.

The power recovery fraction for the binary process, which is independent of the number of stages, is 0.42, the same as for the single-stage flash steam process. The lowest output power is from the flash binary process. The power recovery fraction varies from only 0.18 with one stage to 0.39 with four stages.

The reason for the increase in power recovery fraction with number of stages in the case of the flash steam and flash binary processes is the reduction in brine discharge temperature and, in the case of the flash binary process, the better matching of temperatures on the two sides of the heat exchanger.

Figure 27 compares the power recovery fractions for 150°C reservoir temperature when pumping is used; the flow to the conversion processes is saturated

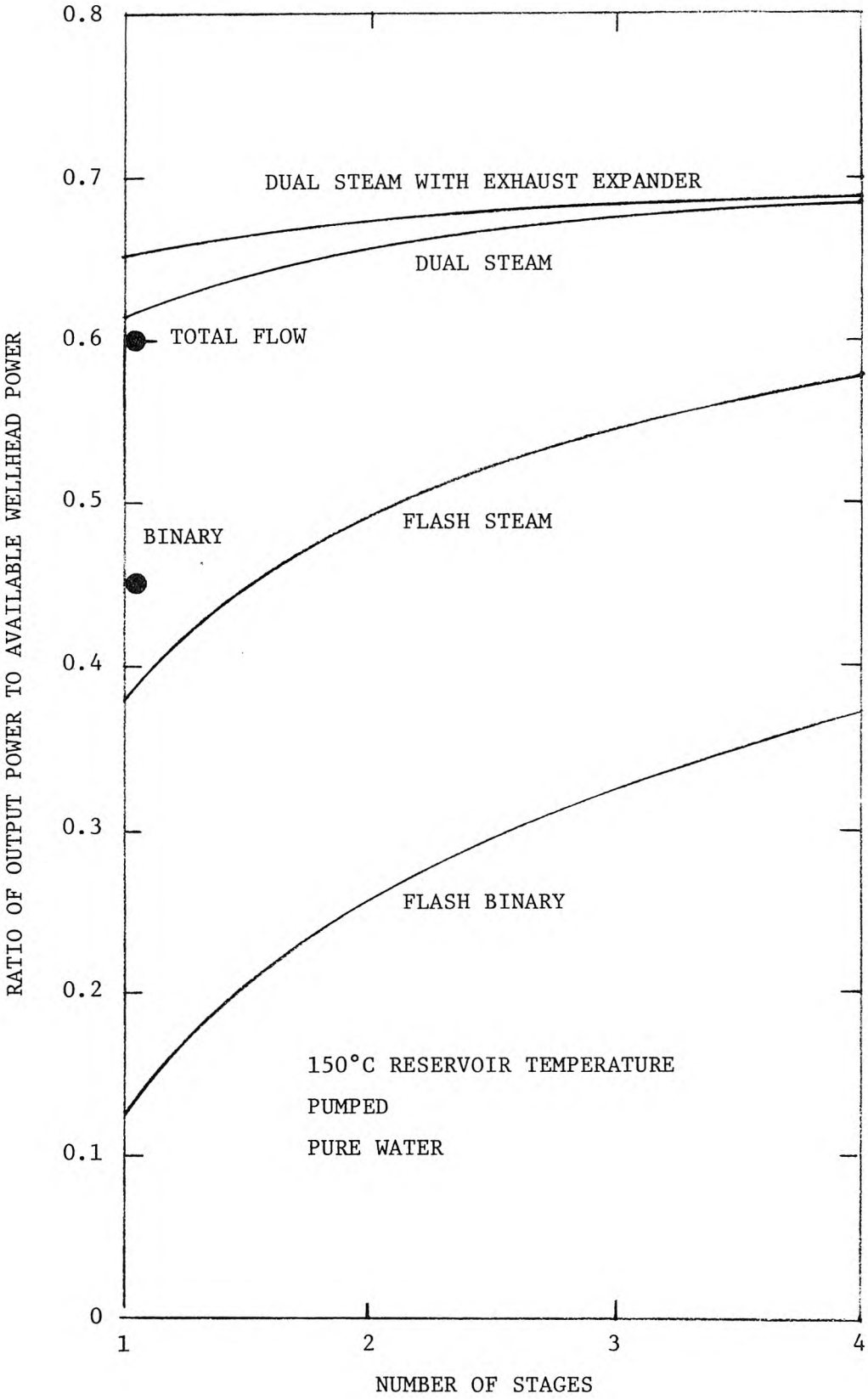

Fig. 27 Comparison of power recovery fractions for 150°C reservoir temperature, pumped, pure water

liquid at 137°C. The dual steam processes give essentially the same power recovery fractions as in the previous case with two-phase wellhead flow. The power recovery fraction for the total flow process is still 0.6. The flash steam process gives about 10 percent less power than with the two-phase wellhead flow. The binary process gives a power recovery fraction of 0.45, and the flash binary ranges from 0.13 to 0.37.

Figure 28 compares the power recovery fractions for self-flowing pure water at 300°C reservoir-temperature. The flow to the conversion processes is water and steam at 218°C and 0.21 quality. Again, the dual steam processes give 0.64 to 0.7 power recovery fraction, and the total flow process gives 0.6. The power recovery fractions of the flash steam process are closer to those of the dual steam and total flow processes at this higher temperature, ranging from 0.47 with one stage to 0.66 with four stages. The power for the binary process is slightly higher than that of the single-stage flash steam process but is lower for two or more stages of flashing. The flash binary process again has the lowest power recovery fraction ranging from 0.29 with one stage to 0.50 with four stages.

Figure 29 compares the conversion processes for self-flowing 30 percent brine at 300°C reservoir temperature; the flow to the conversion processes is brine and steam at 202°C and 0.12 quality. The effect of the high brine concentration is small, the power recovery fractions being essentially the same as for pure water.

To illustrate the heat exchanger temperature distributions in the binary processes, Figure 30 shows how the temperatures vary along the two sides of the heat exchanger for various water inlet conditions at 300°C reservoir temperature with pure water. The bottom curve shows the temperature rise of the isobutane as heat is added. The isobutane pressure is above critical, but the temperature still follows an S-curve analogous to the boiling region at lower pressures. The dashed curve shows the temperatures on the water side with one stage of flashing, and the stair-step curve shows the water-side temperatures with four stages of flashing; the higher final temperature of the isobutane and better matching of the temperatures accounts for the higher power recovery fraction of the four-stage system. Because of the constant water-side temperatures during condensing, the temperature difference is always lowest at the ends of the heat

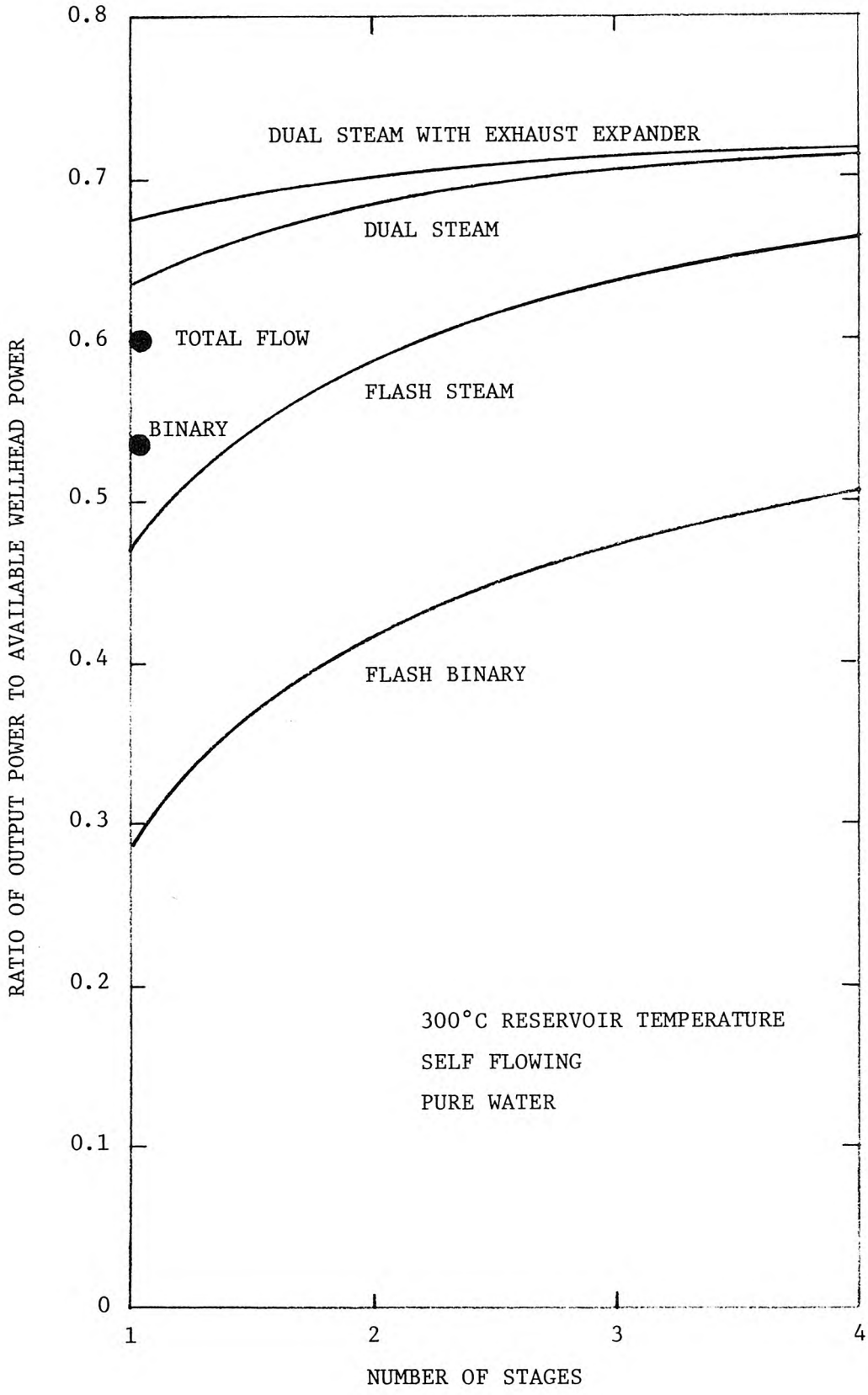

Fig. 28 Comparison of power recovery fractions for 300°C reservoir temperature, self flowing, pure water

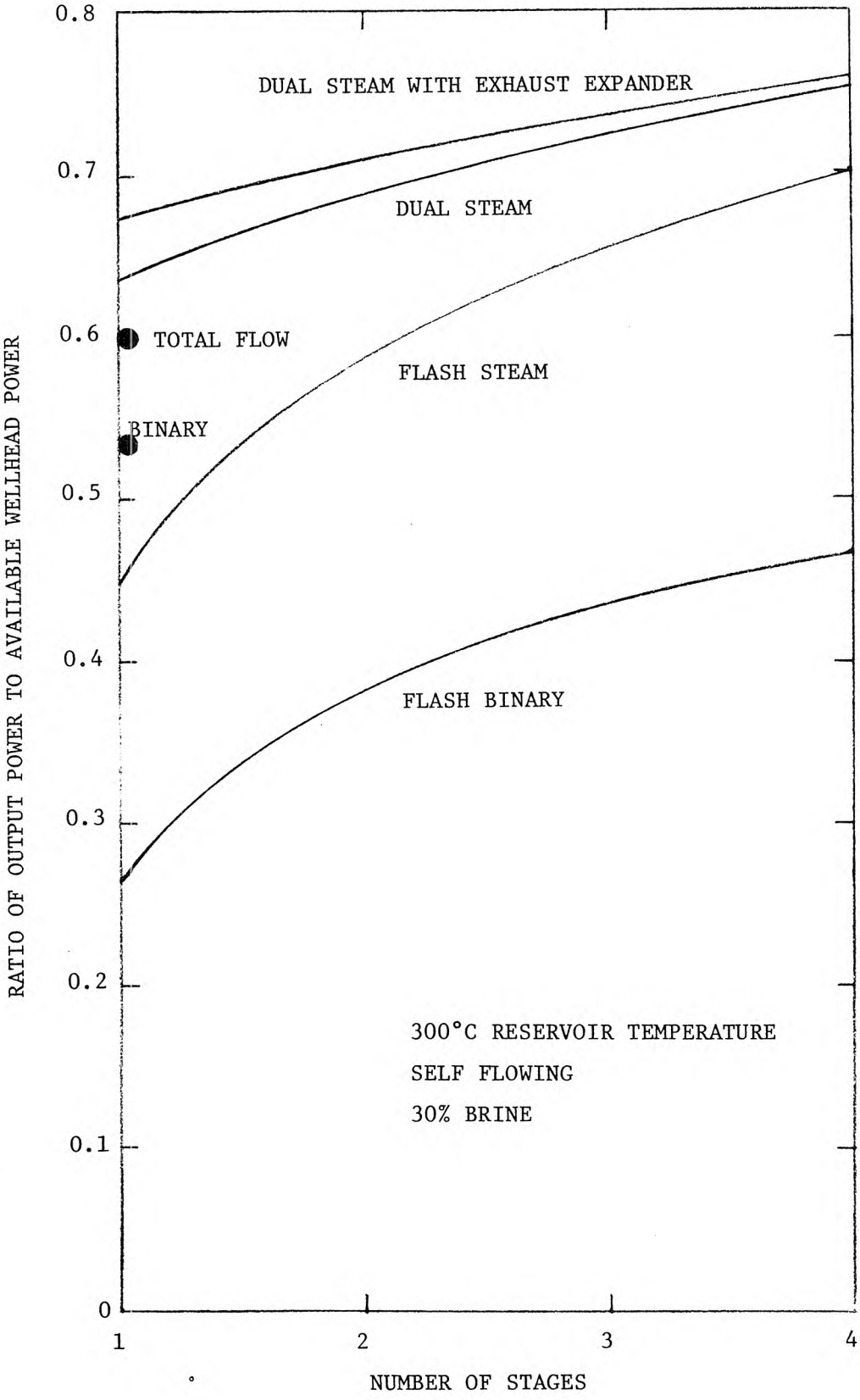

Fig. 29 Comparison of power recovery fractions for 300°C reservoir temperature, self flowing, 30% brine

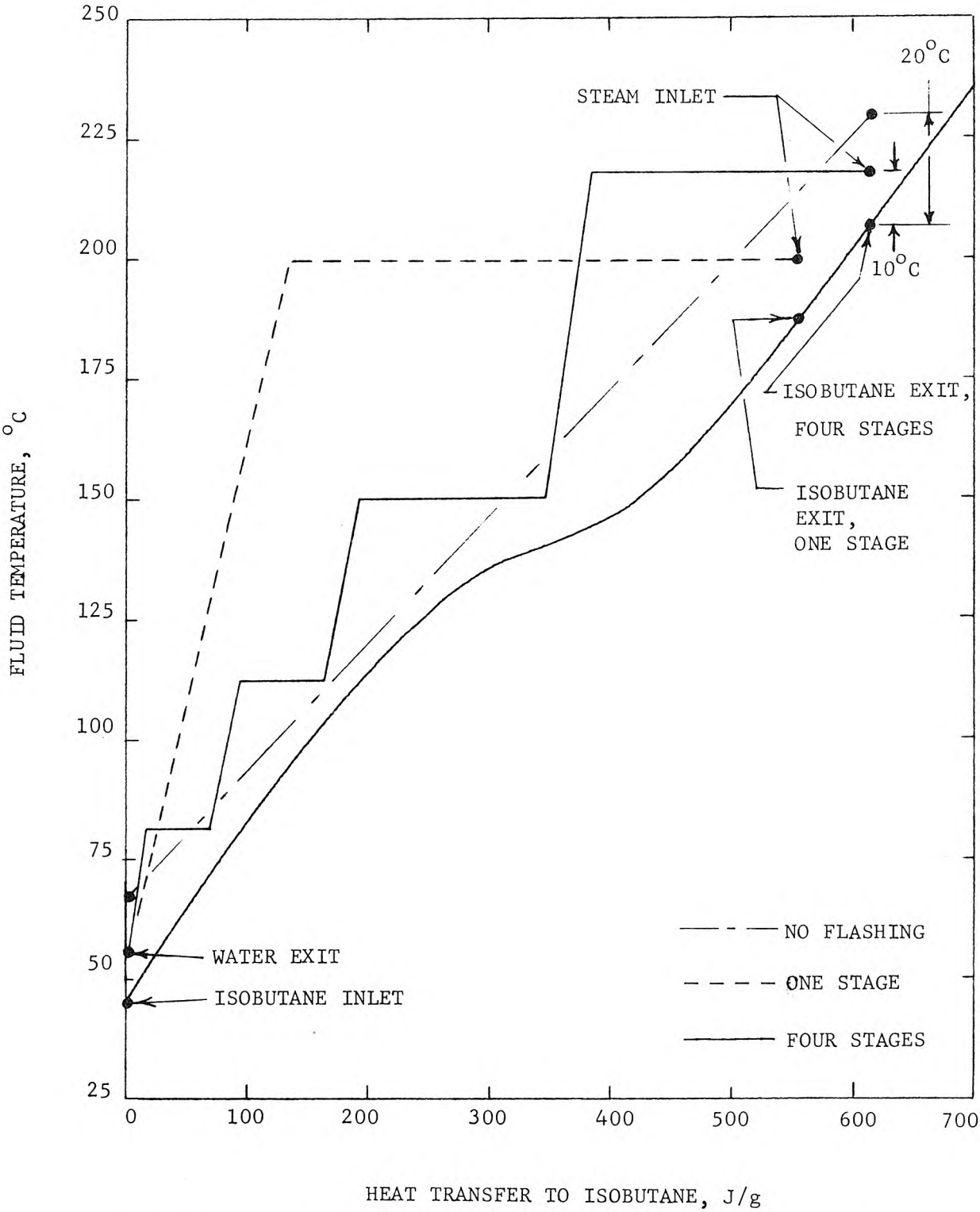

Fig. 30 Heat exchanger temperatures for flash steam binary process with 300°C reservoir temperature, pure water

exchangers and can be set to a low value such as the 10°C specified in the calculations with steam present. If a pumped well is used, however, with no flashing of the water, the water-side temperature follows a straight line from the inlet to the exit and intersects the isobutane curve unless a temperature difference of at least 20°C is specified at the inlet and exit. This is the reason for specifying a larger temperature difference of 20°C for binary with a pumped well (Table 4).

J. Comparison of Condenser Heat Outputs

The amount of power generated per unit of condenser heat rejection is important as a measure of the cooling water consumption and cooling tower cost for given plant size. Figures 31 and 32 compare the ratios of power output to condenser heat rejection for self-flowing pure water at 150°C and 300°C reservoir temperatures, respectively. The flash steam and dual steam processes have ratios of power output to condenser heat rejection of 9-10 percent at the 150°C reservoir temperature and 21-24 percent at the 300°C reservoir temperature. The total flow process has less favorable ratios of 8 percent at 150°C and 18 percent at 300°C. The binary and flash binary processes have still lower ratios of 5-6 percent at 150°C and 15-17 percent at 300°C.

K. Comparison of Steam Condensing Temperatures

If the wellhead flow contains a large amount of noncondensable gases, it may not be possible to use vacuum condensing in the conversion process. At the least, vacuum pumps or steam ejectors will be required, and they will consume power. A comparison of conversion processes with noncondensables present will favor binary processes, because it is possible to transfer heat from the wellhead flow and cool the brine to any desired temperature without reducing the pressure. The conversion processes can be compared on the basis of lowest steam condensing temperature, as a measure of lowest pressure required and of pumping requirement for noncondensables.

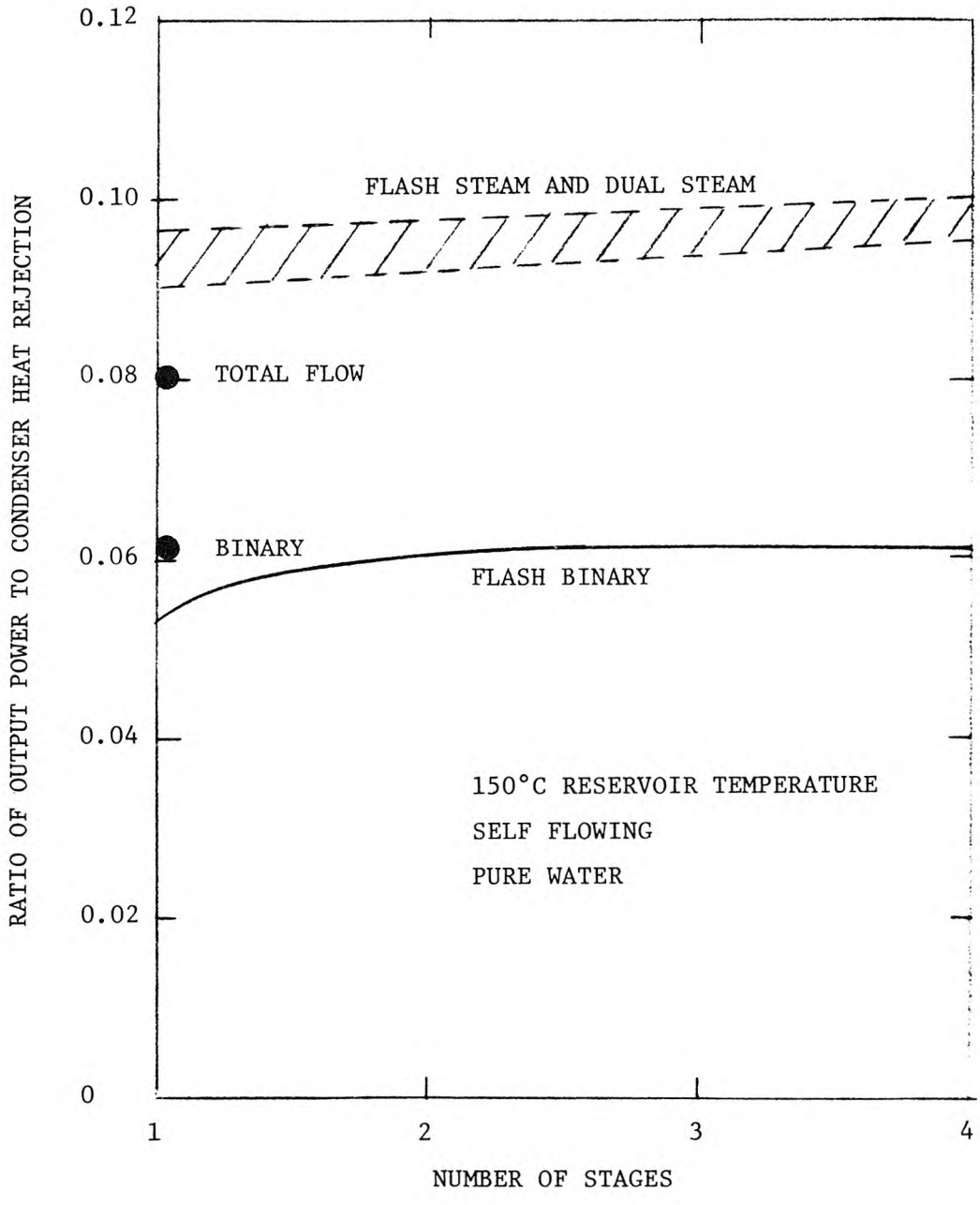

Fig. 31 Comparison of power outputs per unit condenser heat rejection for 150°C reservoir temperature, self flowing, pure water

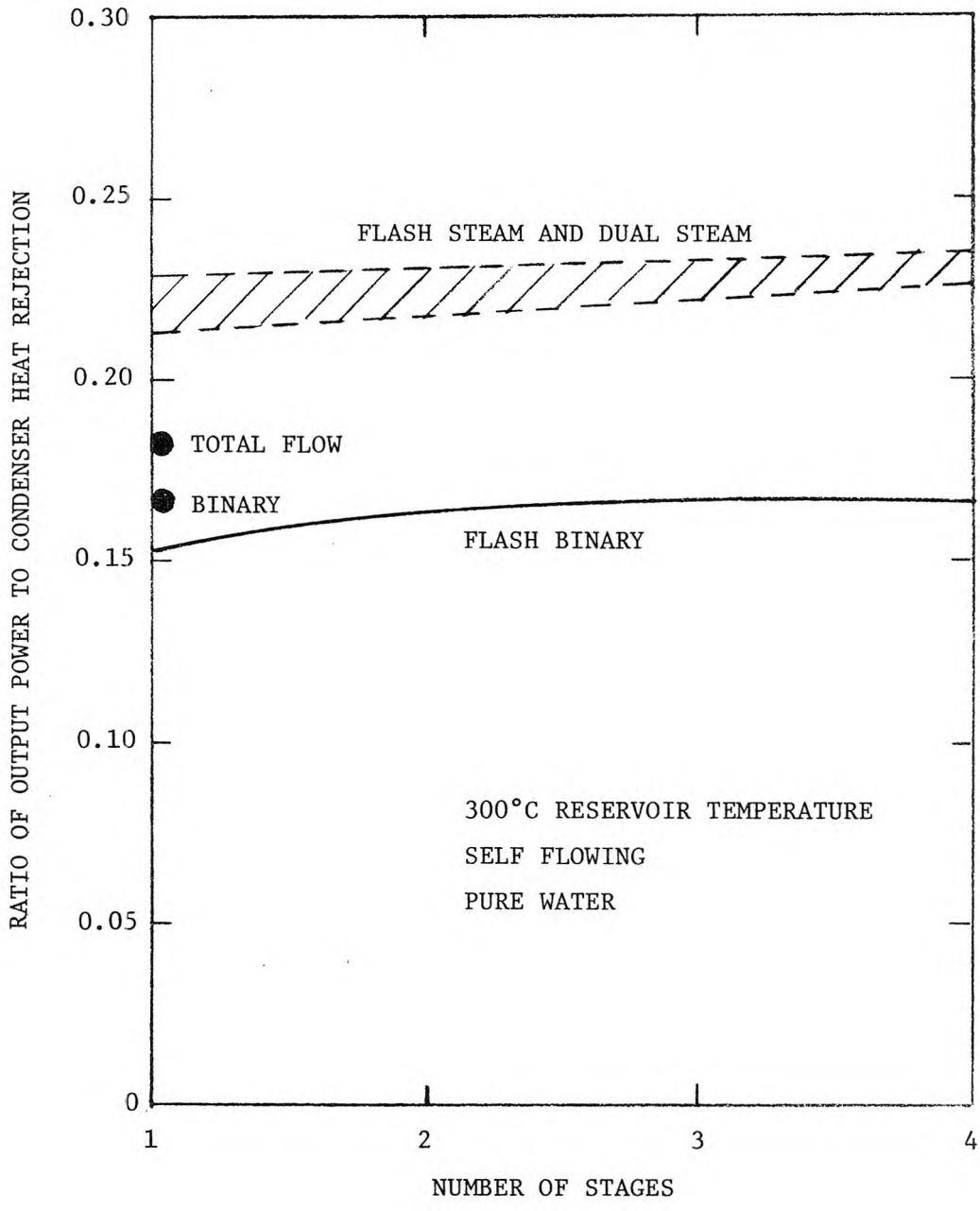

Fig. 32 Comparison of power outputs per unit condenser heat rejection for 300°C reservoir temperature, self flowing, pure water

Figure 33 compares the lowest steam condensing temperatures of the various conversion processes for 150°C reservoir temperature and self flowing. The lowest steam condensing temperature for the binary process is 114°C (the wellhead temperature) and for the flash steam, dual steam, and total flow processes is 45°C (the conversion process rejection temperature). The lowest steam condensing temperature for flash binary is 110°C with one stage but less than 100°C for two stages or more, requiring pumping of noncondensables.

The lowest steam condensing temperatures are compared for 300°C reservoir temperature in Figure 34. Here, flash binary is more favorable with respect to condensing pressure than at the lower temperature, requiring below-atmosphere condensing only with more than three stages.

L. Two-Phase Expander Volume Flows

Two-phase expanders for the dual steam or total flow processes are likely to be more bulky machines for given volume throughout than steam turbines. The velocity of the two-phase flow is less than that of the steam flow in a steam turbine, and the fraction of the machine cross section that can be devoted to flow area may be less than in a steam turbine.

As in the case of steam turbines, by far the largest two-phase expander will be the lowest-pressure one, and the size and cost requirement for two-phase expanders for the processes using them can be compared, on a relative basis, by comparing the volume flow rates from the last-stage two-phase expanders. Figures 35 and 36 present such a comparison for 150°C and 300°C reservoir temperatures, respectively. The volume flow rate leaving the last two-phase expander in the total flow process is 2.6 m^3 per kg of wellhead flow at 150°C and 6.2 m^3/kg at 300°C; there is no change with number of stages. The volume flow rate leaving the last two-phase expander in the dual steam processes is much smaller, only 1 m^3/kg with one stage and 0.4 m^3/kg with four stages at 150°C and 1-2 m^3/kg with one stage and 0.5-0.6 m^3/kg with four stages at 300°C. Compared with the total flow process, the volume flows for a two-stage dual steam process, for example, are only 25 percent as much as 150°C and 10 percent as much at 300°C.

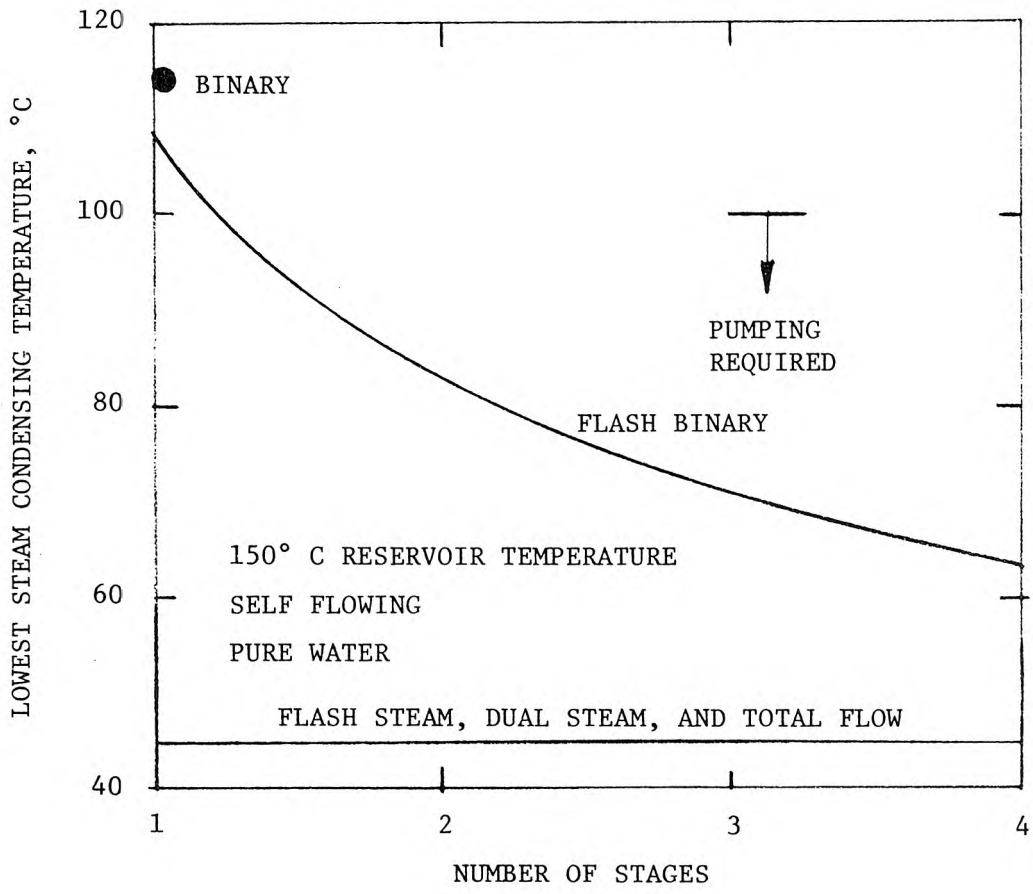

Fig. 33 Comparison of lowest steam condensing temperatures for 150°C reservoir temperature, self flowing, pure water

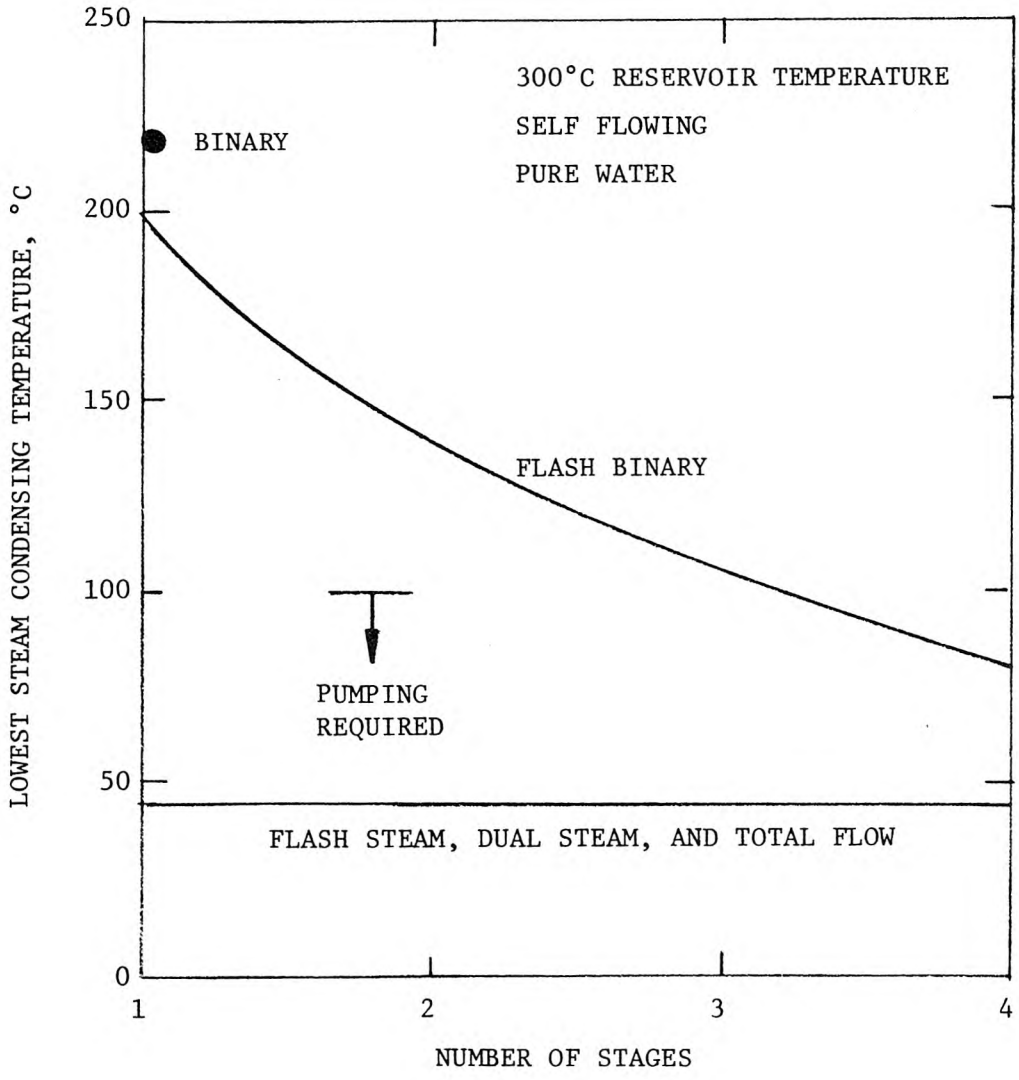

Fig. 34 Comparison of lowest steam condensing temperatures for 300°C reservoir temperature, self flowing, pure water

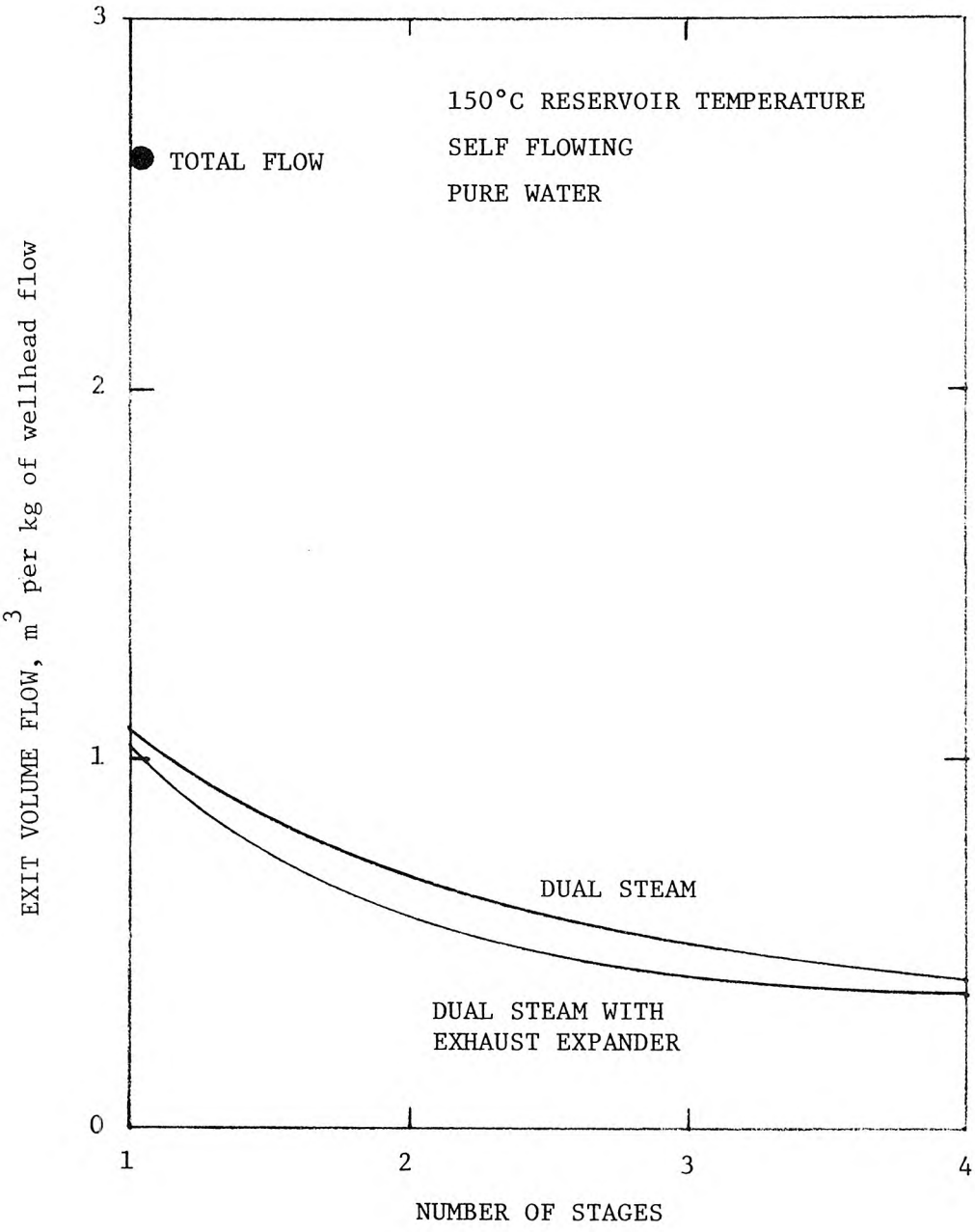

Fig. 35 Exit volume flow from last two-phase expander for 150°C reservoir temperature, self flowing, pure water

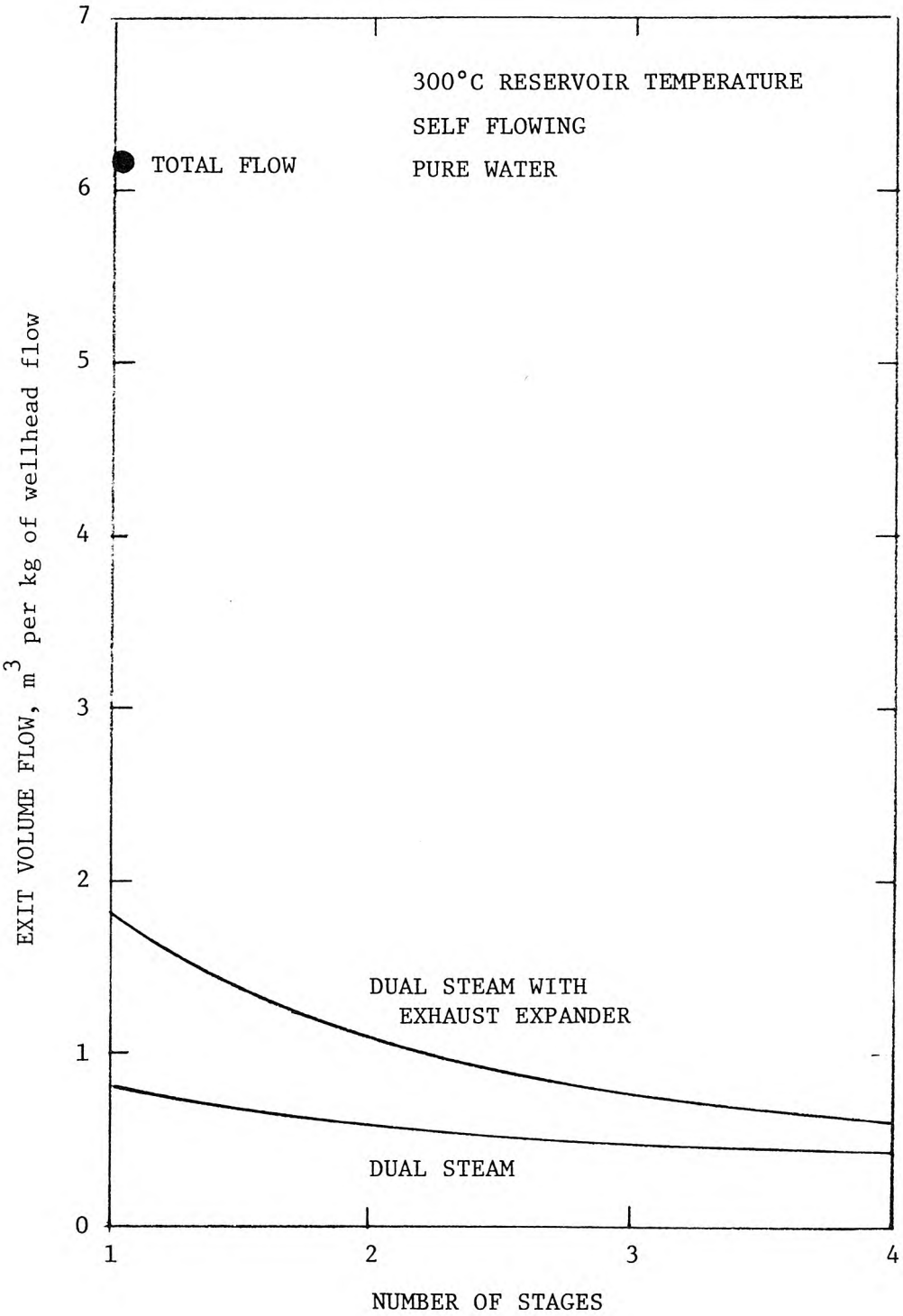

Fig. 36 Exit volume flow from last two-phase expander for 300°C reservoir temperature, self flowing, pure water

The reason for the lower volume flows in the dual steam processes is the fact that most of the steam is handled by the steam turbines. The reason for the decrease in last-stage volume flow with increasing number of stages is that the last stage starts with saturated liquid at a lower temperature. If an exhaust liquid expander is used the exit volume flow may be higher, as in Figure 36, due to the lower expander exit pressure; but the volume flow may be the same or even less, as in Figure 35, if the reoptimization of stage temperatures for maximum power leads to a sufficiently lower inlet temperature to the exhaust expander.

M. Comparison with Other Studies

The results presented here agree with previous calculations for flash steam and flash binary processes[7] and for the binary process.[8] Comparisons with the calculations of Reference 7 are presented in Table 5. The wellhead conditions assumed in Reference 7 correspond to a reservoir temperature of about 320°C. The steam turbine efficiencies assumed are slightly lower than used here and decrease with number of stages because no moisture removal is assumed between stages. Using the constants from Reference 7, the power recovery fractions calculated by the method used here are within 0.03 of the values calculated in Reference 7 for one-, two-, and three-stage flash steam and for two- and three-stage flash binary.

Comparisons with the calculations of Reference 8 are presented in Table 6. Reference 8 presents results for both an isobutane cycle and an advanced cycle denoted as the H-H cycle. The assumptions regarding heat exchanger temperature difference and turbine efficiency are not given, and the values assumed here are the ones in Table 4. The results are higher than the Reference 8 results for isobutane but agree within 0.04 with the results for the H-H cycle.

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(7) A Technology Assessment of Geothermal Resource Development, The Futures Group, Glastonbury, Connecticut, April 1975, pp. 49-50.

(8) Cortez, D. H., Holt, B., and Hutchinson, A. J. L., "Advanced Binary Cycles for Geothermal Power Generation," Energy Sources, Volume 1, No. 1973.

Table 5 Comparison with Flash Steam and Flash Binary Calculations of Reference 7

Wellhead conditions:	219°C temperature 0.175 quality 0.27 brine concentration
Condensing temperature:	38°C (steam) 46°C (isobutane)
Steam turbine efficiency:	0.673 (1 stage) 0.663 (2 stages) 0.657 (3 stages)

Conversion Process	Power Recovery Fraction	
	Reference 7	Present Method
One-stage flash steam	0.49	0.46
Two-stage flash steam	0.55	0.57
Three-stage flash steam	0.62	0.63
Two-stage flash binary	0.37	0.36
Three-stage flash binary	0.41	0.42

Table 6 Comparison with Binary Calculations of Reference 8

Condensing Temperature: 43°C

Wellhead Conditions			Power Recovery Fraction		
Temperature °C	Quality	Brine Conc.	Reference 8 (Table 4)		Present Method
			Isobutane	H-H Process	Isobutane
177	0	0.10	0.66	0.69	0.69
232	0	0.18	0.58	0.71	0.66
182	0.083	0.14	0.48	0.55	0.57
204	0.207	0.30	0.39	0.53	0.50

N. Summary

The relative ranking of the conversion processes with respect to power output is the same for all of the wellhead conditions considered: dual steam with exhaust expander the highest; dual steam without exhaust expander slightly lower; total flow slightly lower than single-stage dual steam; flash steam substantially lower with one stage, but equal to or higher than total flow and within 10 percent of dual steam with three or four stages; binary higher than single-stage flash, but lower than two-stage flash steam; and flash binary lowest, with no more than 50 percent power recovery fraction even with four stages.

The ranking with respect to lowest steam condensing pressure and noncondensable pumping requirement, however, is binary best, with highest pressure and no noncondensable pumping requirement; flash binary next, with condensing pressures ranging from above atmospheric to about 25 percent of atmospheric; and flash steam, dual steam, and total flow worst with condensing pressures about 10 percent of atmospheric.

IV. CONCLUSIONS

The conclusions of this study are:

1. The flow rate from a self-flowing well can be substantially increased by increasing well bore diameter at and above the flashing level.
2. Self-flowing stepped diameter wells provide more wellhead power than pumped wells (for equal production zone diameters) at nearly all reservoir conditions.
3. Multi-stage flash steam systems using steam turbines provide more output power than binary systems at all reservoir temperatures, using reasonable values for component performance.
4. Total flow machines, used alone, offer no performance advantage as compared to multi-stage direct flash steam, unless efficiency of total flow devices can be increased above current projections.

5. Two phase expanders in combination with steam turbines provide the highest output power.

These conclusions derive from theoretical calculations of system performance. They do not include the effects of excessive noncondensable gas content in the reservoir fluid, considerations of scaling and corrosion, or cost and engineering characteristics of the various production and conversion methods. If verified, however, the concept of stepping well diameter may provide a very significant cost saving in the development of a geothermal field. The calculations indicate that flow per well might be increased by a factor of two, thus perhaps halving costs of well drilling for a given level of power.

Detailed engineering studies (see, for example, Reference 9) are now suggesting that binary conversion systems may be quite expensive, owing to the large heat exchangers, pumps and other plumbing required. If the cost of a binary system exceeds that of a direct steam system, and offers no performance advantage, as suggested by these calculations, the binary system may be of value only in very special circumstances.

V. RECOMMENDATIONS

Several reserach and development recommendations are suggested by the conclusions of the study.

1. Perform well flow tests (possibly first at laboratory scale) to verify the predictions for self-flowing with enlarged casing in the upper portion.
2. Carry out engineering studies of construction methods for stepped-diameter production wells.
3. Determine if well plugging (scaling) in flashing flow is serious enough to require pumping instead of self-flowing.
4. If the predicted flow rates with self-flowing are verified, and plugging is not a great problem, reduce priority for development of well pumps.

5. Determine if reservoirs with high noncondensable gas content are a significant proportion of the total resource. If not, reduce priority for development of binary processes.

6. Adopt the multi-stage flash steam process as the basic geothermal power plant.

7. Develop two-phase expanders and add them later to the flash steam plants. If their efficiency can be made sufficiently high, they may become the preferred conversion device.

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E R R A T U M

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for a Proof-of-Concept Facility -- ERDA Grant AER 74-19931 A01
Bechtel Corporation, San Francisco, California, May 1975

APPENDIX A

CALCULATION METHOD FOR SELF-FLOWING WELL

A. Nomenclature

AP	pipe area
C	sonic velocity
CF	skin friction coefficient
CONC	brine concentration
DEPTH	well depth
DFCTR	factor of increase of well diameter above height DFX
DFX	height above bottom at which diameter changes
DH	enthalpy change in interval T + 1 to T
DIA	well diameter (at bottom, if stepped)
DIAX	variable well diameter
DKE	change in kinetic energy per kg in interval T + 1 to T
DMOM	change in momentum in interval T + 1 to T
DP	change in pressure in interval T + 1 to T or between flash level and bottom
DQ	heat transferred out of well in interval T + 1 to T or between the bottom and the flash level
DW	work done per kg by mixture in interval T + 1 to T
DY	height change in interval T + 1 to T
GRAV	acceleration due to gravity, 9.80665 m/s^2
H(T)	enthalpy per kg of mixture at T
HB(T)	enthalpy of saturated brine at T
HGB(T)	enthalpy of vapor in equilibrium with brine at T
HSTAG	enthalpy per kg of flow after stagnation
HTC	heat transfer coefficient for heat flow out of well
HTR	enthalpy per kg of mixture after isentropic expansion to temperature TR
KD	drawdown pressure factor
KE(T)	kinetic energy of mixture per kg at T
KV	ratio of gas velocity to liquid velocity
MACH(T)	Mach number at T

MG	gas (steam) flow rate
ML	liquid (brine) flow rate
MOM(T)	momentum of mixture at T
MW	well flow rate MG + ML
P	pressure
PAV(T)	specific available power at wellhead for mixture at temperature T
PRES	reservoir pressure at well bottom
PSATB(T)	saturation pressure of brine at T
PW	wellhead pressure
P1	well bottom pressure
RA	ratio of gas flow area to liquid flow area
RHOB(T)	density of brine at T
RHOGB(T)	density of vapor in equilibrium with brine at T
RHOL(T)	density of pure water at T
RHOM(T)	two-phase mixture density at T
RHOMM	mean two-phase mixture density over interval T + 1 to T
RM	ratio of gas flow rate to liquid flow rate MG/ML
S(T)	entropy per kg of mixture at T
SB(T)	entropy of saturated brine at T
SGB(T)	entropy of vapor in equilibrium with brine at T
SSTAG	entropy per kg of mixture after stagnation
T	temperature
TR	rejection temperature for calculating available power
TRES	reservoir temperature
TSINK	sink temperature for well heat loss
TW	wellhead temperature
T1	well bottom temperature
VG	gas (steam) velocity
VL(T)	liquid (brine) velocity at T
WS	wall shear
X(T)	quality MG/MW at T
XSTAG	quality after stagnation
XTR	quality after isentropic expansion to temperature TR
Y	height above well bottom

B. Assumptions

The assumption used in the analysis of self flowing are:

1. The flow is steady.
2. There is no gain or loss in flow rate between the well bottom and wellhead.
3. The liquid is incompressible.
4. In the two-phase region the phases are in equilibrium.
5. The phases are homogeneously mixed.
6. The velocity of each phase is constant across the well pipe.
7. The ratio of gas velocity to liquid velocity is constant.
8. The flow leaving the wellhead is stagnated at constant pressure.
9. The temperature outside the well varies linearly from 20°C at ground level to the reservoir temperature at the bottom.
10. Any stepped diameter starts at or above the flash level.

Assumption 1 enables steady-flow relations to be used and corresponds to normal operation of a geothermal well. Assumption 2 is met if the bottom of the well is taken to be the location of the highest perforation or inlet and there is no leakage out of the well.

Assumption 3 permits use of saturation liquid density for compressed liquid and introduces negligible error. Assumption 4 permits use of saturation thermodynamic properties and is accurate because of the relatively slow changes in temperature and pressure in the mixture flowing up the well.

Assumption 5 allows wall shear to be calculated for homogeneous flow and is accurate because geothermal well flow is in the homogeneous two-phase flow regime. Assumption 6 permits flow rate, momentum, and kinetic energy to be calculated directly from the velocities and is valid because the flow is turbulent with a relatively flat velocity profile. Assumption 7 is adopted to allow study of flow with slip between the phases under the simplification of constant velocity ratio, but slip is apparently negligible, as shown later, and the velocity ratio is not only constant but unity.

Assumption 8 is in accord with the usual method of handling the wellhead flow by stagnating it with no attempt to recover the kinetic energy in a diffuser or turbine.

Assumption 9 is a rough approximation to the heat sink temperature outside the well for heat transfer calculations. Assumption 10 is a permissible restriction since there is little gain from enlarging the diameter below the flash level.

C. Reservoir Conditions

The brine concentration CONC and brine temperature TRES at the bottom of the well are specified. The value of CONC is used in the brine property calculations of Appendix C. The zero-flow reservoir pressure PRES at the bottom of the well is either specified or calculated from the hydrostatic pressure of ground water:

$$\text{PRES} = \text{RHOL}(20)\ \text{DEPTH}\ \text{GRAV} \tag{A-1}$$

where RHOL(20) is the density of pure water at 20°C, DEPTH is the specified well depth, and GRAV is the acceleration due to gravity.

For calculating the specific available power of the brine in the reservoir by the procedure derived in Subsection G, the height is set to Y = 0, the temperature to T = TRES, and the kinetic energy to KE(T) = 0. The enthalpy is the enthalpy of saturated brine HB(T) plus the work required to pump the liquid from saturation pressure PSATB(T) to reservoir pressure PRES.

$$H(T) = HB(T) + \frac{\text{PRES} - \text{PSATB}(T)}{\text{RHOB}(T)} \tag{A-2}$$

where the density of the brine is approximated by the saturation value PSATB(T) in accordance with Assumption 3.

D. Well Bottom Conditions

The well bottom temperature T1 is the same as the reservoir temperature. The well flow rate MW is specified, and the well bottom pressure P1 is calculated

from the specified drawdown pressure factor KD.

$$P1 = PRES - KD\ MW \tag{A-3}$$

The enthalpy is the enthalpy of saturated brine plus the work to pump the liquid from saturation pressure to P1.

$$H(T) = HB(T) + \frac{P1 - PSATB(T)}{RHOB(T)} \tag{A-4}$$

The specified pipe, or casing, diameter is DIA. If the well has stepped diameter the diameter is increased to DIAX = DFCTR DIA above height DFX and remains at DIAX = DIA below DFX. The pipe area is

$$AP = \frac{\pi}{4} DIAX^2 \tag{A-5}$$

The liquid velocity is

$$VL(T) = \frac{MW}{RHOB(T)\ AP} \tag{A-6}$$

The kinetic energy per unit mass is

$$KE(T) = \frac{1}{2} VL(T)^2 \tag{A-7}$$

E. Flash-Level Conditions

In accordance with Assumption 10, the diameter and velocity are constant between the well bottom and flash level. The wall shear for the liquid flowing upward at velocity VL(T) is

$$WS = CF \frac{RHOB(T)}{2} VL(T)^2 \tag{A-8}$$

where CF is the specified skin friction coefficient.

In accordance with Assumption 9 that the heat sink temperature varies linearly from 20°C at the top of the well to TRES at the bottom, the mean heat sink temperature between the well bottom and height Y, the flash level, is

$$\text{TSINK} = \text{TRES} - \frac{(\text{TRES} - 20)\ Y}{2\ \text{DEPTH}} \tag{A-9}$$

The heat transferred from the well per unit flow rate between the bottom and the flash level is

$$\text{DQ} = \frac{\text{HTC}\ \pi\ \text{DIA}\ Y\ (T - \text{TSINK})}{\text{MW}} \tag{A-10}$$

where HTC is the specified heat transfer coefficient.

The flash-level temperature T is the temperature at which the brine enthalpy has been decreased by amount DQ from well-bottom enthalpy.

$$\text{HB}(T) = \text{HB}(T1) - \text{DQ} \tag{A-11}$$

The pressure at the flash level is the saturation pressure PSATB(T). The pressure difference between the flash level and well bottom is

$$\text{DP} = \text{PSATB}(T) - P1 \tag{A-12}$$

The downward forces acting on the column of liquid between the flash level and the bottom are pressure force DP AP, weight RHOB(T) AP Y GRAV, and shear force WS π DIA Y. Setting the sum of those forces to zero, the flash-level height is

$$Y = \frac{-\ \text{DP}\ \text{AP}}{\text{RHOB}(T)\ \text{AP}\ \text{GRAV} + \text{WS}\ \pi\ \text{DIA}} \tag{A-13}$$

If the value of Y calculated from Eq. (A-13) is greater than the well depth this means that the pressure at the wellhead is greater than saturation and flashing does not occur. The wellhead conditions can then be calculated immediately

by interpolation. The wellhead temperature is

$$TW = T1 + \frac{DEPTH}{Y} (T - T1) \tag{A-14}$$

and the wellhead pressure is

$$PW = P1 + \frac{DEPTH}{Y} DP \tag{A-15}$$

F. Two-Phase Region

In the two-phase region between the flash level and the top the calculations must be performed in small increments of temperature and height so that average values of varying quantities such as density can be used. Temperature increments of 1°C are sufficiently small. Starting at the flash-level temperature, the temperature is decreased 1°C at a time and the changes in flow conditions between the old temperature T + 1 and the new temperature T are calculated.

First, an approximation to the quality X(T) at the new temperature T is calculated based on an isentropic change.

$$X(T) = \frac{S(T + 1) - SB(T)}{SGB(T) - SB(T)} \tag{A-16}$$

where S(T + 1) is the previously-calculated entropy of the mixture at T + 1, SB(T) is the entropy of saturated brine at T, and SGB(T) is the entropy of vapor in equilibrium with brine at T.

The pressure difference from the previous temperature is

$$DP = PSATB(T) - PSATB(T + 1) \tag{A-17}$$

The gas-phase flow rate is

$$MG = X(T)\ MW \tag{A-18}$$

and the liquid flow rate is the difference between the total well flow and the gas flow.

$$ML = MW - MG \tag{A-19}$$

The liquid velocity is VL(T) and the gas velocity is, in general, a different value

$$VG = KV\ VL \tag{A-20}$$

where KV is the specified ratio of gas velocity to liquid velocity. The liquid flow area is ML/[RHOB(T) VL(T)] and the gas flow area is MG/[RHOGB(T) KV VL(T)] where RHOGB(T) is the density of vapor in equilibrium with the brine. Equating the sum of the liquid and gas flow areas to the pipe area AP the liquid velocity is

$$VL(T) = \frac{1}{AP}\left(\frac{ML}{RHOB(T)} + \frac{MG}{RHOGB(T)\ KV}\right) \tag{A-21}$$

The mass of liquid in unit height of pipe is ML/VL and the mass of gas is MG/VG. The sum of the two is equal to the total mass RHOM AP in unit height of pipe, where RHOM is the density of the two-phase mixture. Thus, the mixture density is

$$RHOM(T) = \frac{1}{AP}\left(\frac{ML}{VL(T)} + \frac{MG}{VG}\right) \tag{A-22}$$

The mean mixture density over the interval T + 1 to T is

$$RHOMM = \frac{1}{2}\left[RHOM(T) + RHOM(T+1)\right] \tag{A-23}$$

The momentum of the mixture is

$$MOM(T) = ML\ VL(T) + MG\ VG \tag{A-24}$$

and the change in momentum over the interval T + 1 to T is

$$DMOM = MOM(T) - MOM(T + 1) \tag{A-25}$$

The kinetic energy per unit mass of mixture is

$$KE(T) = \frac{ML\ VL^2 + MG\ VG^2}{2\ MW} \tag{A-26}$$

and the change in kinetic energy is

$$DKE = KE(T) - KE(T + 1) \tag{A-27}$$

The wall shear of a two-phase mixture is primarily due to the liquid phase and is equal to the shear that would exist with liquid alone at the same velocity, multiplied by the fraction of the wall that is wetted.[A-1] Under Assumption 5 that the mixture is homogeneous, the fraction of the wall that is wetted is the ratio of mixture density RHOM to liquid density. Thus, the wall shear is simply CF RHOM $VL^2/2$, and the mean wall shear over the interval T + 1 to T is

$$WS = \frac{CF}{4}\left|RHOM(T)\ VL(T)^2 + RHOM(T + 1)\ VL(T + 1)^2\right| \tag{A-28}$$

The downward forces acting on the column of mixture of height DY between T + 1 and T are pressure force DP AP, weight RHOMM AP DY GRAV, shear force WS π DIAX DY, and momentum increase DMOM. Setting the sum of those forces to zero, the height change between T + 1 and T is

$$DY = \frac{-\ DP\ AP - DMOM}{RHOMM\ AP\ GRAV + WS\ \pi\ DIAX} \tag{A-29}$$

A-1 Eddington, R. B., "Investigation of Supersonic Shock Phenomena in a Two-Phase (Liquid-Gas) Tunnel," JPL Technical Report No. 32-1096, March 15, 1967, p. 181.

The work DW done by unit mass of mixture over the interval T + 1 to T is the work of lifting the mass a distance DY plus the increase in kinetic energy. Thus,

$$DW = GRAV\ DY + DKE \tag{A-30}$$

The average heat sink temperature for heat transfer over the interval Y to Y + DY is

$$TSINK = TRES - (TRES - 20)\ \frac{Y + DY/2}{DEPTH} \tag{A-31}$$

and the heat transfer out of the well, per unit mass, over the height change DY is

$$DQ = \frac{HTC\ \pi\ DIAX\ DY\ (T - TSINK)}{MW} \tag{A-32}$$

By the First Law, the enthalpy change over the interval T + 1 to T is

$$DH = - DQ - DW \tag{A-33}$$

Hence, the enthalpy of the mixture at the new temperature T is

$$H(T) = H(T + 1) + DH \tag{A-34}$$

The corresponding quality is

$$X(T) = \frac{H(T) - HB(T)}{HGB(T) - HB(T)} \tag{A-35}$$

This value of X replaces the value calculated from Eq. (A-16), and the procedure from Eq. (A-18) through Eq. (A-35) is repeated until X(T) converges to the correct value.

The temperature is then lowered another 1°C and the next set of flow conditions is calculated. The procedure ends in one of two ways. If the height change

puts the new height above the wellhead, the wellhead conditions are found by interpolation, ending the calculation. If a negative DY is found this means that the flow cannot reach the wellhead because the flow rate exceeds the value for choking (the flow can only reach the level at which the DY reversal occurs).

G. Available Wellhead Power

The power available at the wellhead from 1 kg/s of brine or brine and steam mixture at any height Y in the well is the enthalpy difference for an isentropic expansion of 1 kg of the mixture to rejection temperature TR, minus the work to raise the mixture against gravity to the wellhead elevation. The initial conditions are temperature T, mixture enthalpy H(T) per kg, mixture kinetic energy KE(T) per kg, and height Y. According to Assumption 8, the flow is first stagnated at constant pressure, raising the enthalpy to the stagnation enthalpy

$$HSTAG = H(T) + KE(T) \tag{A-36}$$

The quality XSTAG after stagnation is zero for compressed liquid (and negligibly different from zero for saturated liquid). For initial two-phase flow the quality after stagnation is such that the enthalpy is HSTAG = (1 - XSTAG) HB(T) + XSTAG HGB(T), where HB(T) is the enthalpy of saturated brine at temperature T and HGB(T) is the enthalpy of steam in equilibrium with the brine. Thus, the quality after stagnation is

$$XSTAG = \frac{HSTAG - HB(T)}{HGB(T) - HB(T)} \tag{A-37}$$

The entropy of the 1 kg mixture after stagnation is

$$SSTAG = (1 - XSTAG)\ SB(T) + XSTAG\ SGB(T) \tag{A-38}$$

where SB(T) is the entropy of saturated brine at temperature T and SGB(T) is the entropy of steam in equilibrium with the brine.

An isentropic expansion to rejection temperature TR yields a two-phase mixture of quality XTR such that the entropy is SSTAG = (1 - XTR) SB(TR) + XTR SGB(TR),

where SB(TR) and SGB(TR) are the entropies of the brine and steam, respecively, at temperature TR. Thus, the quality after the isentropic expansion is

$$XTR = \frac{SSTAG - SB(TR)}{SGB(TR) - SB(TR)} \tag{A-39}$$

The enthalpy of the 1 kg mixture after the isentropic expansion is

$$HTR = (1 - XTR)\ HB(TR) + XTR\ HGB(TR) \tag{A-40}$$

where HB(TR) and HGB(TR) are the enthalpies of the brine and steam, respectively, at temperature TR.

The work available is the enthalpy difference HSTAG - HTR for the isentropic expansion. If the mixture is at height Y above the bottom of the well the available work at the wellhead is HSTAG - HTR minus the work to raise 1 kg the distance DEPTH - Y. Thus, the specific available wellhead power in the mixture at height Y is

$$PAV(T) = HSTAG - HTR - GRAV\ (DEPTH - Y) \tag{A-41}$$

H. Mach Number

The sonic velocity in a two-phase mixture, from Reference A-1, p. 3, is

$$C = (1 + RA)\left[\frac{PSATB(T)}{RHOB(T)\ RA\ (1 + RM)}\right]^{1/2} \tag{A-42}$$

where RA is the ratio of gas flow area MG/[RHOGB(T) VG] to liquid flow area ML/[RHOB(T) VL(T)] and RM is the ratio of gas flow rate MG to liquid flow rate ML. The ratio of flow areas RA is used in place of the volume ratio employed in Reference A-1, because RA is the appropriate measure of the liquid volume fraction when there is slip between the phases and is the same as the volume ratio in the absence of slip. The Mach number of the flow, defined on the basis of liquid

velocity, is

$$MACH(T) = \frac{VL(T)}{C}$$

I. Comparison with Experiment and Choice of CF and KV

Measurements for the No. 1 IID well in the Salton Sea area are presented by Helgeson.[A-2] The well constants inferred from information in Reference A-1 are: temperature TRES = 295°C, well depth DEPTH = 1520 m, well diameter DIA = 0.178 m, brine concentration CONC = 0.3, and drawdown pressure factor KD = 22.8 kPa per kg/s. The reservoir pressure at the well bottom is that due to normal hydrostatic gradient (Eq. (A-1)), namely, PRES = 14.9 MPa (2160 psi).

Figure 10 of Reference A-2 presents steam and brine flow rates as a function of wellhead pressure from choked flow to about half of choked flow. The measured total well flow rate is reproduced here in Figure A-1 and compared to the theory for three combinations of skin friction coefficient CF and gas-to-liquid velocity ratio KV. The best agreement is for CF = 0.008 and KV = 1.0, although a slightly lower CF would improve the agreement at high flow rates. Increasing the gas-to-liquid velocity ratio has little effect at high flow rates but decreases the wellhead pressure at low flow rates. Decreasing the skin friction coefficient raises the flow rate at low pressures but has little effect at high pressures and low flow rates.

Figure A-2 compares the theoretical and measured steam and liquid flow rates for CF = 0.008 and KV = 1.0. The main differences are higher measured values of steam flow rate and zero-flow pressure than predicted by the theory.

J. Computer Program

A Fortran IV computer program incorporating the equations derived in this Appendix is presented on the following pages. Results for a test case are also given.

A-2 Helgeson, H.C., "Geologic and Thermodynamic Characteristics of the Salton Sea Geothermal System," American Journal of Science, Vol. 266, pp. 129-166, March 1968.

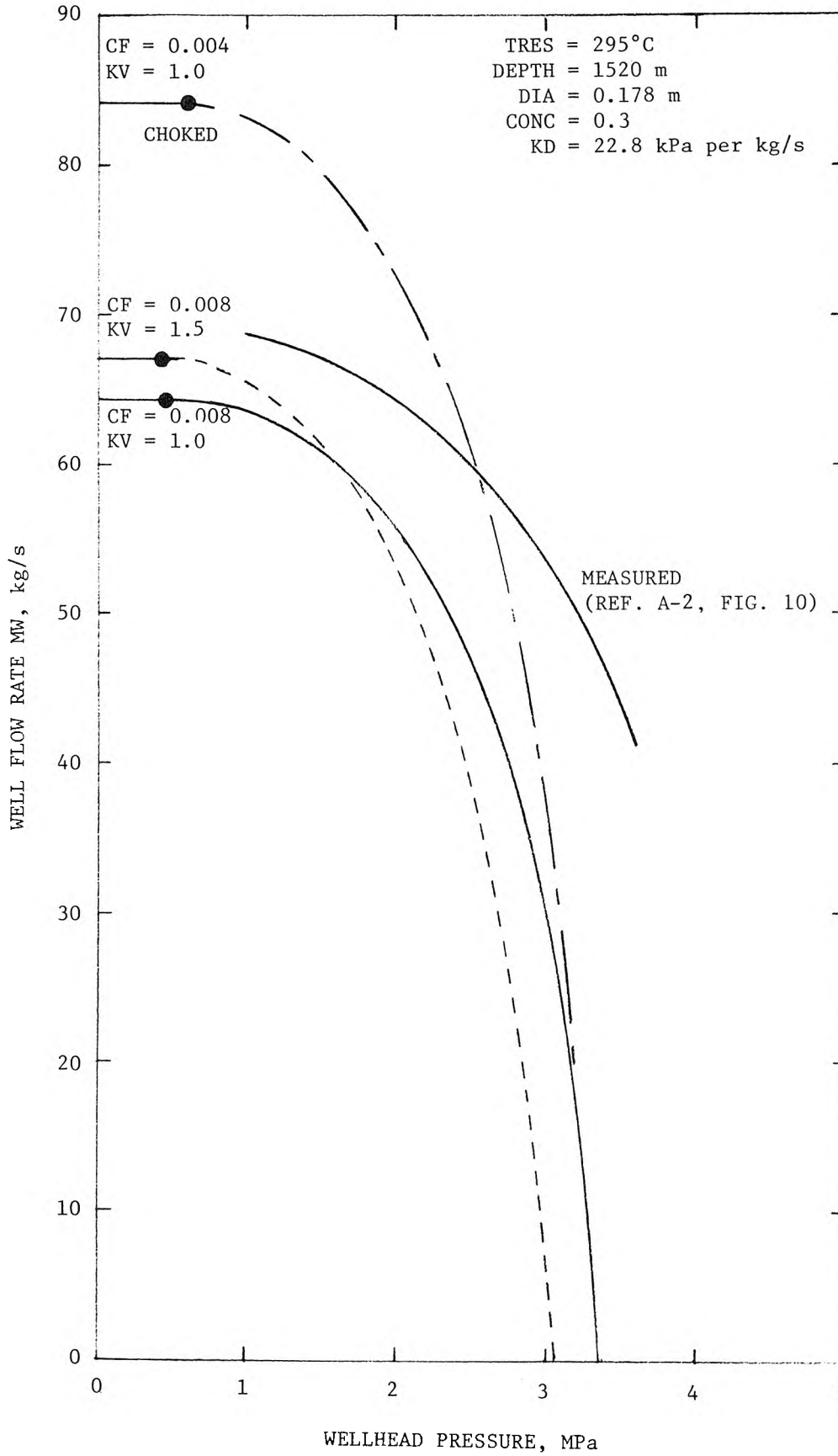

Fig. A-1 Comparison between theoretical and measured total flow rates in No. 1 IID well with three combinations of skin friction coefficient CF and gas/liquid velocity ratio KV

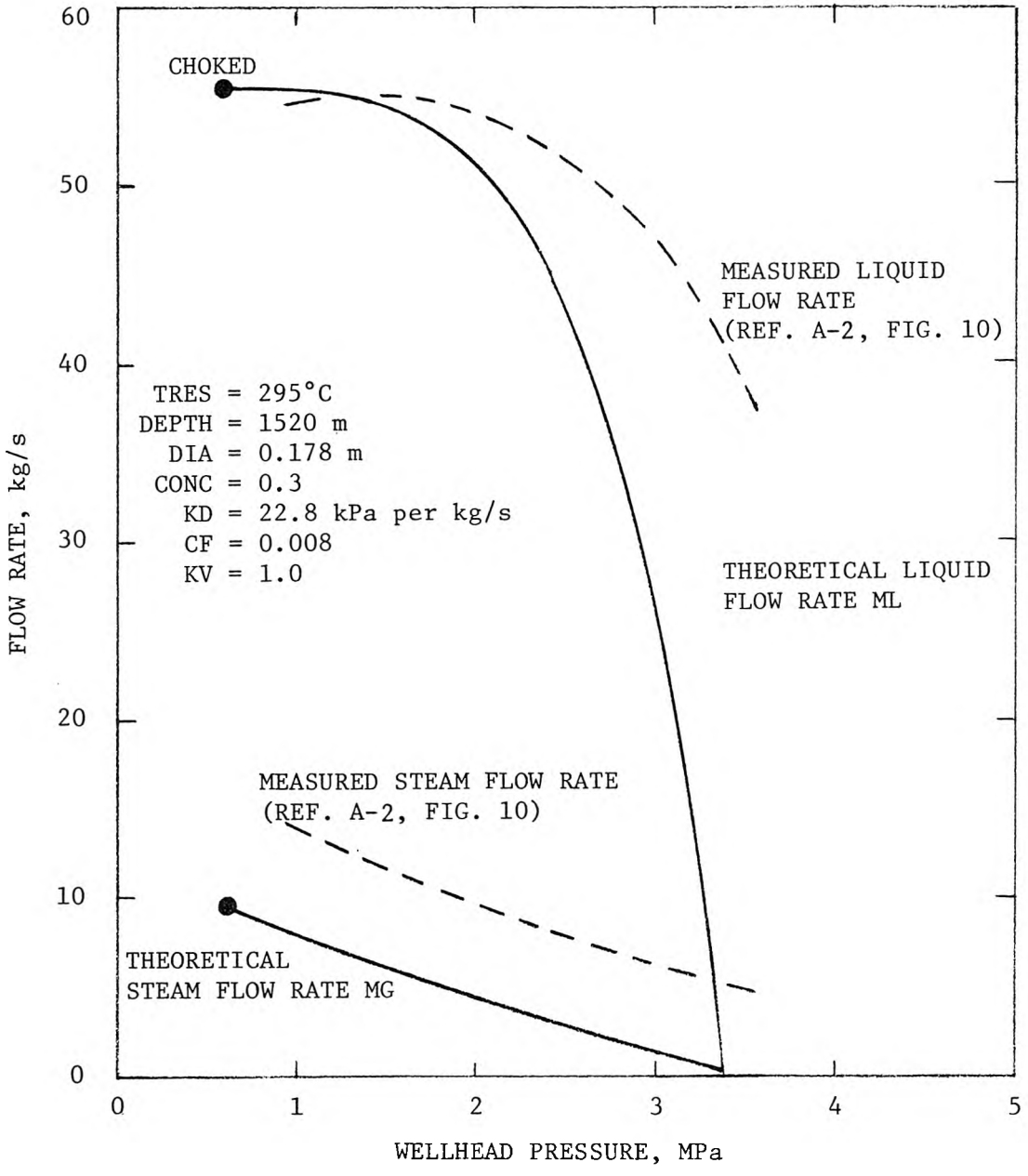

Fig. A-2 Comparison between theoretical and measured steam and liquid flow rates in No. 1 IID well

```
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C  * * * * * * * * * * SELF-FLOWING GEOTHERMAL WELL  * * * * * * * * * *
C
C  * * * * * * * * * * D. G. ELLIOTT, MARCH 24, 1975 * * * * * * * * * *
C
C  * * * * * * * * * *  LAST REVISION, AUG. 5, 1975  * * * * * * * * * *
C
C  CALCULATES THE WELLHEAD CONDITIONS FOR A SELF-FLOWING GEOTHERMAL
C  WELL GIVEN THE DEPTH, RESERVOIR TEMPERATURE AND PRESSURE, BRINE
C  CONCENTRATION, WELL DIAMETER, AND FLOW RATE.
C
C  * * * * * * * * EQUATION NUMBERS REFER TO APPENDIX A  * * * * * * * *
C
C  INPUT DATA
C  ----------
C
C  CONC     BRINE CONCENTRATION FRACTION (DENVER RESEARCH INSTITUTE MIX)
C  TRES     RESERVOIR TEMPERATURE, DEG C
C  PRES     RESERVOIR PRESSURE, PA (SEE NOTE 1)
C  MW       WELL FLOW RATE, KG/S (SEE NOTE 2)
C  DIA      WELL DIAMETER, M
C  DEPTH    WELL DEPTH, M
C  DFCTR    FACTOR OF INCREASE OF WELL DIAMETER ABOVE DFX (SEE NOTE 3)
C  DFX      HEIGHT ABOVE BOTTOM AT WHICH DIAMETER INCREASES (SEE NOTE 3)
C  KD       DRAWDOWN PRESSURE FACTOR, PA PER KG/S
C  KV       RATIO OF GAS-PHASE VELOCITY TO LIQUID VELOCITY
C  CF       SKIN FRICTION COEFFICIENT
C  HTC      HEAT TRANSFER COEFFICIENT, J/M*2 PER DEG C (SEE NOTE 4)
C  TR       REJECTION TEMPERATURE FOR AVAILABLE POWER EVALUATION
C  NPR      '0' FOR NOTE-1 RESERVOIR PRESSURE, '1' FOR INPUT
C  NINPT    '1' FOR ECHO OF INPUTS
C  NDET     '1' FOR PRINTOUT OF WELL CONDITIONS AT 1 DEG STEPS
C  NSRCH    '1' FOR PRINTOUT OF TOTAL AVAILABLE POWER ONLY
C
C  OUTPUTS
C  -------
C
C  WELL CONDITIONS:
C
C  Y, M          HEIGHT ABOVE WELL BOTTOM, METERS
C  P, MPA        PRESSURE, MEGAPASCALS
C  T, C          TEMPERATURE, DEG C
C  X             QUALITY
C  VL, M/S       LIQUID VELOCITY, M/S
C  MACH NO       MACH NUMBER
C  AVAIL, J/G    AVAILABLE POWER, JOULES PER GRAM (KW PER KG/S)
C
C  THE FIRST LINE OF PRINTOUT GIVES THE RESERVOIR CONDITIONS.
C  THE SECOND LINE GIVES THE WELL BOTTOM CONDITIONS.
C  THE THIRD LINE GIVES THE FLASH-LEVEL CONDITIONS.
C  THE LAST LINE GIVES THE WELLHEAD CONDITIONS.
C  IF NDET=1, INTERVENING LINES GIVE CONDITIONS BETWEEN FLASH AND TOP.
C
C  TOTALS:
C
```

```
C   MW              WELL FLOW RATE, KG/S
C   TOTAL AVAIL     TOTAL AVAILABLE WELLHEAD POWER, MW
C
C   NOTE 1: IF NPR=0, RESERVOIR PRESS IS SET EQUAL TO THE NORMAL HYDRO-
C           STATIC VALUE = COLUMN OF 20 DEG C PURE WATER OF WELL DEPTH.
C
C   NOTE 2: IF THE SPECIFIED FLOW RATE EXCEEDS THE CHOKED FLOW RATE FOR
C           THE WELL, THE CONDITIONS AT THE DEPTH WHERE CHOKING WOULD
C           OCCUR AT THAT FLOW RATE ARE PRINTED.
C
C   NOTE 3: OMITTING DFCTR AND DFX GIVES CONSTANT DIAMETER
C
C   NOTE 4: SINK TEMP ASSUMED LINEAR FROM 20 C AT TOP TO TRES AT BOTTOM.
C
      PARAMETER MXT=300 @ MAX TEMP IN PROPERTY TABLES, CAN USE LESS
C
      INTEGER T,TRES,TR,TM1,SHORT
      IMPLICIT REAL(M)
      REAL PSAT(MXT),HL(MXT),HG(MXT),SL(MXT),SG(MXT),RHOL(MXT),
     &     RHOG(MXT),PSATB(MXT),HB(MXT),HGB(MXT),SB(MXT),SGB(MXT),
     &     RHOB(MXT),RHOGB(MXT),X(MXT),S(MXT),H(MXT),VL(MXT),
     &     RHOM(MXT),MOM(MXT),KE(MXT),MACH(MXT),PAV(MXT),KV,KD
C
C     EQUIVALENCE (PSAT,PSATB),(HL,HB),(HG,HGB),(SL,SB),(SG,SGB),
C    &            (RHOL,RHOB),(RHOG,RHOGB) @ FOR PURE WATER CASES
C
      NAMELIST /IN/CONC,TRES,PRES,MW,DIA,DEPTH,DFCTR,DFX,KD,KV,
     &           CF,HTC,TR,NPR,NINPT,NDET,SHORT
C
C   ***** LOAD PURE WATER THERMODYNAMIC PROPERTIES *****
C
      CALL H2OPRP(PSAT, HL, HG, SL, SG, RHOL, RHOG,
     &            PSATB, HB, HGB, SB, SGB, RHOB, RHOGB, MXT)
C
C   ***** DEFAULT INPUTS AND CONSTANTS *****
C
      KD=25.0E3 @ DRAWDOWN PRESSURE OF IID NO.1 (REF. HELGESON)
      DIA=0.25 @ 25-CM CASING DIAMETER.
      KV=1.0 @ GAS/LIQUID VELOCITY RATIO
      CF=0.008 @SKIN FRICTION COEFFICIENT
      TR=45 @ REJECTION TEMPERATURE, DEG C
      PI=3.14159
      GRAV=9.80665 @ ACCELERATION DUE TO GRAVITY, M/S**2
      Z=0 @ FOR PRINTING ZERO
C
C ***** INPUTS *****
C
10    PRINT 15
15    FORMAT (//' ENTER')
      READ (5,IN,ERR=10,END=95) @ READ INPUTS
      IF (NPR.EQ.0) PRES=RHOL(20)*DEPTH*GRAV @NORMAL RESERVOIR PRESS (1)
      IF (NINPT.EQ.1) WRITE (6,IN) @ ECHO INPUTS
C
C   ***** LOAD BRINE THERMODYNAMIC PROPERTIES *****
C
      OLDCON=-1 @ TO ASSURE CALL ON FIRST PASS
      IF (CONC.NE.OLDCON) CALL BRNPRP(CONC) @ LOAD BRINE PROPERTIES
```

```
      OLDCON=CONC @ TO PREVENT REPEAT CALL IF CONC NOT CHANGED
C
C   ***** RESERVOIR CONDITIONS *****
C
      Y=0 @ RESERVOIR HEIGHT IS ZERO
      T=TRES @ RESERVOIR TEMP IS TRES
      KE(T)=0 @ RESERVOIR KINETIC ENERGY IS ZERO
      H(T)=HB(T)+(PRES-PSATB(T))/RHOB(T) @COMPRESSED LIQUID ENTHALPY (2)
      NCL=1 @ TO GIVE COMPRESSED-LIQUID OPTION IN SUBROUTINE AVAIL
      CALL AVAIL @ SET PAV(T)=SPECIFIC AVAILABLE WELLHD PWR IN RESERVOIR
      TT=TRES @ FOR PRINTOUT
      IF (SHORT.EQ.0.AND.NDET.EQ.0) PRINT 30 @ HEADING W/O PAGE EJECT
      IF (SHORT.EQ.0.AND.NDET.EQ.1) PRINT 31 @ HEADING WITH PAGE EJECT
30    FORMAT (/'     Y',7X'P',7X'T',7X'X',7X'VL      MACH    AVAIL',/
     &         '     M',6X'MPA',6X'C',14X'M/S       NO      J/G'/)
31    FORMAT ( '1    Y',7X'P',7X'T',7X'X',7X'VL      MACH    AVAIL',/
     &         '     M',6X'MPA',6X'C',14X'M/S       NO      J/G'/)
      IF (SHORT.EQ.0) PRINT 35,Y,PRES,TT,Z,Z,Z,PAV(T) @ RES CONDITIONS
35    FORMAT (F7.1,-6PF8.2,0PF8.1,F8.4,F8.1,F8.4,-3PF8.1)
C
C   ***** WELL BOTTOM CONDITIONS *****
C
      T1=TRES @ NO TEMP CHANGE
      P1=PRES - KD*MW @ RESERVOIR PRESS MINUS DRAWDOWN (3)
      H(T)=HB(T)+(P1-PSATB(T))/RHOB(T) @ COMPRESSED LIQUID ENTHALPY (4)
      DIAX=DIA @ DIAX IS SPECIFIED DIAMETER BELOW STEP
      AP=PI/4 *DIAX**2 @ PIPE AREA (5)
      VL(T)=MW/(RHOB(T)*AP) @ WELL BOTTOM VELOCITY (6)
      KE(T)=VL(T)**2 /2 @ KINETIC ENERGY PER KG (7)
      CALL AVAIL @ SET PAV(T) = POWER AVAILABLE AT BOTTOM
      IF (SHORT.EQ.0) PRINT 35,Y,P1,T1,Z,VL(T),Z,PAV(T) @ BOTTOM COND
C
C   **** FLASH-LEVEL CONDITIONS (OR WELLHEAD CONDITIONS IF NO FLASH) ****
C
      WS=CF*RHOB(T)/2 *VL(T)**2 @ WALL SHEAR GUESS (8)
      DP=PSATB(T)-P1 @ PRESS DIFF GUESS BET FLASH LEVEL AND BOTTOM (12)
      Y=-DP*AP/(RHOB(T)*AP*GRAV+WS*PI*DIA) @ FLASH HEIGHT GUESS (13)
      TSINK=TRES - (TRES-20)*Y/(2*DEPTH) @ AVERAGE HEAT SINK TEMP (9)
      DQ=HTC*PI*DIA*Y*(T-TSINK)/MW @HEAT LOSS PER KG TO FLASH LEVEL (10)
C
          DO 40 T=TRES,20,-1 @ FIND TEMP AT FLASH LEVEL TO NEAREST DEG
40        IF (HB(T).LT.HB(TRES)-DQ) GO TO 45 @T IS BETWEEN T & T+1 (11)
C
45    T=T+1 @ SET T TO NEXT HIGHER INTEGRAL DEG C
      DP=PSATB(T) - P1 @ PRESS DIFF BETWEEN FLASH LEVEL AND BOTTOM (12)
      VL(T)=MW/(RHOB(T)*AP) @ LIQUID VELOCITY BEFORE FLASH (6)
      WS=CF*RHOB(T)/2 *VL(T)**2 @ WALL SHEAR BEFORE FLASH (8)
      Y=-DP*AP/(RHOB(T)*AP*GRAV + WS*PI*DIA) @ FLASH HEIGHT (13)
      H(T)=HB(T) @ ENTHALPY OF LIQUID BEFORE FLASH
      KE(T)=VL(T)**2 /2 @ KINETIC ENERGY BEFORE FLASH (7)
      NCL=0 @ TO GIVE TWO-PHASE OPTION IN AVAIL
      CALL AVAIL @ SET PAV(T) = POWER AVAILABLE AT FLASH LEVEL
      TT=T @ FLASH-LEVEL TEMPERATURE
      PP=PSATB(T) @ FLASH-LEVEL PRESSURE
      PV=PAV(T) @ FLASH-LEVEL AVAILABLE POWER
      IF (Y.GT.DEPTH) TT=T1 + DEPTH*(T-T1)/Y @ WELLHEAD TEMPERATURE (14)
      IF (Y.GT.DEPTH) PP=P1 + DEPTH*DP/Y @ WELLHEAD PRESSURE (15)
```

```
      IF (Y.GT.DEPTH) PV=PAV(TRES)+DEPTH*(PAV(T)-PAV(TRES))/Y @ WLHD PWR
      IF (Y.GT.DEPTH) Y=DEPTH @ REDUCE Y TO WELLHEAD HEIGHT
      IF (SHORT.EQ.0) PRINT 35,Y,PP,TT,Z,VL(T),Z,PV @ FLASH OR TOP
      IF (Y.GE.DEPTH) GO TO 10 @ ASK FOR NEW INPUTS
C
C  ***** TRAVERSE OF TWO-PHASE REGION TO TOP *****
C
      S(T)=SB(T) @ INITIAL ENTROPY IS BRINE ENTROPY
      X(T)=0 @ INITIAL QUALITY IS ZERO
      RHOM(T)=RHOB(T) @ INTIAL MIXTURE DENSITY IS BRINE DENSITY
      MOM(T)=MW*VL(T) @ INITIAL MOMENTUM IS LIQUID MOMENTUM
C
            TM1=T-1
            DO 70 T=TM1,20,-1 @ DECREASE TEMPERATURE IN 1 DEG C STEPS
            TT=T @ REAL T FOR PRINTOUT
            X(T)=(S(T+1) - SB(T))/(SGB(T) - SB(T)) @ QUALITY GUESS (16)
            DP=PSATB(T) - PSATB(T+1) @ PRESS CHANGE FROM FROM PREV T (17)
C
                  DO 50 K=1,100 @ ITERATE (UP TO 100 TIMES) FOR Y CHANGE
                  XSAVE=X(T) @ SAVE OLD QUALITY VALUE
                  MG=X(T)*MW @ GAS FLOW RATE (18)
                  ML=MW - MG @ LIQUID FLOW RATE (19)
                  VL(T)=(ML/RHOB(T) + MG/(RHOGB(T)*KV))/AP @ LIQ VEL (21)
                  VG=KV*VL(T) @ GAS VELOCITY (20)
                  RHOM(T)=(ML/VL(T) + MG/VG)/AP @ MIXTURE DENSITY (22)
                  RHOMM=(RHOM(T) + RHOM(T+1))/2 @MEAN DENSITY OVER DY (23)
                  MOM(T)=ML*VL(T) + MG*VG @ MIXTURE MOMENTUM (24)
                  DMOM=MOM(T) - MOM(T+1) @ MOMENTUM INCREASE IN DY (25)
                  KE(T)=(ML*VL(T)**2 + MG*VG**2)/(2*MW) @ MIX K.E./KG (26)
                  DKE=KE(T) - KE(T+1) @ KINETIC ENERGY INCREASE IN DY (27)
                  WS=CF/4*(RHOM(T)*VL(T)**2+RHOM(T+1)*VL(T+1)**2) @SHR(28)
                  DY=(-DP*AP-DMOM)/(RHOMM*AP*GRAV+WS*PI*DIAX)@HT DIFF (29)
                  DW=GRAV*DY + DKE @ WORK PER KG IN HEIGHT CHANGE DY (30)
                  TSINK=TRES-(TRES-20)*(Y+DY/2)/DEPTH @HEAT SINK TEMP (31)
                  DQ=HTC*PI*DIAX*DY*(T-TSINK)/MW @HEAT X PER KG IN DY (32)
                  DH=-DQ-DW @ ENTHALPY CHANGE FROM FIRST LAW (33)
                  H(T)=H(T+1) + DH @ NEW ENTHALPY VALUE (34)
                  X(T)=(H(T)-HB(T))/(HGB(T)-HB(T)) @NEW QUALITY VALUE (35)
50                IF (ABS(X(T)-XSAVE).LT.1E-6) GO TO 60 @ CONVERGED
C
            PRINT 55 @ USED UP 100 ITERATIONS
55          FORMAT (' DIDN''T CONVERGE ON DY')
            GO TO 10 @ ASK FOR NEW INPUTS
60          IF (DY.GT.0) GO TO 68 @ NORMAL TERMINATION WITHOUT CHOKING
C
C  CHOKED-FLOW TERMINATION
C
            PRINT 65 @ MESSAGE THAT FLOW IS CHOKED
65          FORMAT (' FLOW IS CHOKED.')
            TP1=T+1
            IF (SHORT.EQ.0) PRINT 35,Y,PSATB(T+1),TP1,X(T+1),
     &                  VL(T+1),MACH(T+1),PAV(T+1) @ CHOKE-LEVEL CONDITIONS
            GO TO 10 @ ASK FOR NEW INPUTS
C
C  FLOW CONDITIONS AT NEW HEIGHT
C
68          Y=Y+DY @ NEW HEIGHT
```

```
            IF (DFCTR.GT.0.AND.Y.GE.DFX) DIAX=DIA*DFCTR @ LARGER DIA
            IF (DFCTR.GT.0.AND.Y.GE.DFX) AP=PI*DIAX**2/4 @ LARGER AREA
            CALL AVAIL @ SET PAV(T) = PWR AVAILABLE AT Y
            CALL SONIC @ SET MACH(T) = MACH NUMBER AT Y
            IF (NDET.EQ.1.AND.MOD(TRES-T,48).EQ.0) PRINT 31 @ 48 LN/PAGE
            IF (Y.GT.DEPTH) GO TO 75 @ PASSED TOP, INTERPOLATE BACK
70          IF (NDET.EQ.1) PRINT 35,Y,PSATB(T),TT,X(T),VL(T),
     &                   MACH(T),PAV(T) @ WELL CONDITIONS AT NEW HEIGHT
      GO TO 10 @ ASK FOR NEW INPUTS
C
C  ***** INTERPOLATED WELLHEAD CONDITIONS *****
C
75    XY=(DEPTH - (Y-DY))/DY @ FRACTION OF DY INCREMENT TO TOP
      PW=PSATB(T+1) + (PSATB(T)-PSATB(T+1))*XY @ WELLHEAD PRESSURE
      TW=T+1-XY @ WELLHEAD TEMPERATURE
      XW=X(T+1) + (X(T)-X(T+1))*XY @ WELLHEAD QUALITY
      VLW=VL(T+1) + (VL(T)-VL(T+1))*XY @ WELLHEAD LIQUID VELOCITY
      MACHW=MACH(T+1) + (MACH(T)-MACH(T+1))*XW @ WELLHEAD MACH NUMBER
      PAVW=PAV(T+1) + (PAV(T)-PAV(T+1))*XY @ WELLHEAD AVAILABLE PWR
      PAVWT=PAVW*MW @ TOTAL AVAILABLE POWER AT WELLHEAD
      IF (SHORT.EQ.0) PRINT 35,DEPTH,PW,TW,XW,VLW,MACHW,PAVW @ WELLHEAD
      PRINT 80,MW,PAVWT
80    FORMAT (/' MW =',F6.1,' KG/S,  TOTAL AVAIL =',-6PF7.3,' MW')
      GO TO 10 @ ASK FOR NEW INPUTS
95    STOP
C
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
      SUBROUTINE AVAIL @ CALCULATES POWER AVAILABLE PER KG/S AT SURFACE
C     WITH STAGNATION OF FLOW AT CONSTANT PRESSURE, ISENTROPIC EXPAN-
C     SION TO REJECTION TEMPERATURE TR, AND DEDUCTION OF LIFTING WORK.
C
      HSTAG=H(T) + KE(T) @ ENTHALPY AFTER STAGNATION (36)
      XSTAG=(HSTAG-HB(T))/(HGB(T)-HB(T)) @ QUALITY AFTER STAGNATION (37)
      IF (NCL.GT.0) XSTAG=0 @ COMPRESSED LIQUID REGION
      SSTAG=(1-XSTAG)*SB(T)+XSTAG*SGB(T) @ ENTROPY AFTER STAGNATION (38)
      XTR=(SSTAG-SB(TR))/(SGB(TR)-SB(TR)) @ISENTROPIC QUALITY AT TR (39)
      HTR=(1-XTR)*HB(TR)+XTR*HGB(TR) @ ISENTROPIC ENTHALPY AT TR (40)
      PAV(T)=HSTAG-HTR-GRAV*(DEPTH-Y) @ SPEC AVAILABLE WELLHEAD PWR (41)
      RETURN
C
      SUBROUTINE SONIC @ CALCULATES THE SONIC VELOCITY AND MACH NUMBER
C     GIVEN THE GAS AND LIQUID FLOW RATES, GAS AND LIQUID VELOCITIES,
C     GAS AND LIQUID DENSITIES, AND PRESSURE.
C
C  REF: R. B. EDDINGTON, 'INVESTIGATION OF SUPERSONIC SHOCK PHENOMENA
C  IN A TWO-PHASE (LIQUID-GAS) TUNNEL', JPL TECHNICAL REPORT NO.
C  32-1096, MARCH 15, 1967, P. 3.
C
      AG=MG/(RHOGB(T)*VG) @ GAS FLOW AREA
      AL=ML/(RHOB(T)*VL(T)) @ LIQUID FLOW AREA
      RA=AG/AL @ GAS/LIQUID FLOW AREA RATIO
      RM=MG/ML @ GAS/LIQUID MASS RATIO
      C=(1+RA)*SQRT(PSATB(T)/(RHOB(T)*RA*(1+RM))) @ SONIC VELOCITY (42)
      MACH(T)=VL(T)/C @ MACH NUMBER OF LIQUID FLOW (43)
      RETURN
      END
```

TEST CASE

```
$IN CONC = 0.2, TRES = 250, MW - 110, DEPTH = 1500, DFCTR = 2, DFX = 900 $END


ENTER
$IN
CONC     =      .20000000E+00
TRES     =                +250
PRES     =      .14683544E+08
MW       =      .11000000E+03
DIA      =      .25000000E+00
DEPTH    =      .15000000E+04
DFCTR    =      .20000000E+01
DFX      =      .90000000E+03
KD       =      .25000000E+05
KV       =      .10000000E+01
CF       =      .80000000E-02
HTC      =      .00000000E+00
TR       =                +45
NPR      =                 +0
NINPT    =                 +1
NDET     =                 +0
SHORT    =                 +0
$END
```

Y M	P MPA	T C	X	VL M/S	MACH NO	AVAIL J/G
.0	14.68	250.0	.0000	.0	.0000	133.4
.0	11.93	250.0	.0000	2.3	.0000	130.5
868.9	3.42	250.0	.0000	2.3	.0000	130.2
1500.0	1.53	206.8	.0531	4.4	.0374	129.6

MW = 110.0 KG/S, TOTAL AVAIL = 14.260 MW

APPENDIX B

CALCULATION METHOD FOR PUMPED WELL

A. Nomenclature

AP	pipe area
CF	skin friction coefficient
CONC	brine concentration
DEPTH	well depth
DIA	well diameter
DPPUMP	pump pressure rise
DPUMP	pump depth
DTPUMP	brine temperature drop due to heat transfer to the pump turbine loop
DTWELL	brine temperature drop due to well heat loss
GRAV	acceleration due to gravity, 9.80775 m/s^2
HB(T)	enthalpy of saturated brine at T
HTC	heat transfer coefficient for heat loss from well
KD	drawdown pressure factor
KT	brine temperature drop per unit of pump pressure rise due to heat transfer to the pump turbine loop
MW	well flow rate
PAV	specific available wellhead power
PAVT	total available wellhead power
PIN	pump inlet pressure
PPUMP	pumping power
PRES	reservoir pressure at well bottom
PSATB(T)	brine saturation pressure at T
PW	wellhead pressure
P1	well bottom pressure
RHOB(T)	brine density at T
RHOL(T)	density of pure water at T
SB(T)	brine entropy at T
T	temperature
TBAR	mean brine temperature in well
TR	rejection temperature for calculating available power
TRES	reservoir temperature
TSINK	sink temperature for well heat loss

TW	wellhead temperature
T1	well bottom temperature
VL	liquid (brine) velocity
YPUMP	height of pump above well bottom

B. Assumptions

The assumption used in the analysis of pumped well flow are:

1. The flow is steady.
2. There is no gain or loss in flow rate between the well bottom and wellhead.
3. The liquid is incompressible.
4. The pump depth is such that the inlet is at a pressure equal to saturation pressure at the well bottom temperature.
5. The brine is cooled in a heat exchanger immediately above the pump, and the brine temperature drop is proportional to the pump pressure rise.
6. The wellhead flow is saturated liquid.
7. The temperature outside the well varies linearly from $20^{\circ}C$ at ground level to the reservoir temperature at the bottom.

Assumption 1 enables steady-flow relations to be used and corresponds to normal operation of a geothermal well. Assumption 2 is met if the bottom of the well is taken to be the location of the highest perforation or inlet and there is no leakage out of the well. Assumption 3 permits use of saturation liquid density for compressed liquid and introduces negligible error.

Assumption 4 gives the minimum depth at which the pump must be located. The actual depth must be several meters lower to provide a net positive suction pressure, but the effect on the flow calculations is negligible. If there is heat transfer out of the well the pump depth could be less, but this is not taken into account. Assumption 5 represents a pump driven by a vapor turbine using heat transferred from the brine in a heat exchanger attached to the pump. The combined efficiency of the pump, the turbine, and the power loop driving the turbine can be represented by a fixed brine temperature drop per unit pressure use since the ratio of pumping power to heat input is (flow rate x pressure rise)/(flow rate x temperature drop).

Assumption 6 represents the wellhead condition usually assumed for pumping: all liquid with pressure at or slightly above saturation. Assumption 7 is a rough approximation to the heat sink temperature outside the well for heat transfer calculations.

C. Reservoir Conditions

The brine concentration CONC and brine temperature TRES at the bottom of the well are specified. The value of CONC is used in the brine property calculations of Appendix C. The zero-flow reservoir pressure PRES at the bottom of the well is either specified or calculated from the hydrostatic pressure of ground water:

$$\text{PRES} = \text{RHOL}(20)\ \text{DEPTH}\ \text{GRAV} \tag{B-1}$$

where RHOL(20) is the density of pure water at 20°C, DEPTH is the specified well depth, and GRAV is the acceleration due to gravity.

D. Well Bottom Conditions

The well bottom temperature T1 is the same as the reservoir temperature. The well flow rate MW is specified, and the well bottom pressure P1 is calculated from the drawdown pressure factor KD.

$$\text{P1} = \text{PRES} - \text{KD}\ \text{MW} \tag{B-2}$$

The well pipe area is

$$\text{AP} = \frac{\pi}{4}\ \text{DIA}^2 \tag{B-3}$$

where DIA is the well diameter.

The liquid velocity is

$$\text{VL} = \frac{\text{MW}}{\text{RHOB(T1)}\ \text{AP}} \tag{B-4}$$

E. Pump Depth

In accordance with Assumption 4, the pump inlet pressure PIN is equal to the saturation pressure of brine at the well bottom temperature T1.

$$\text{PIN} = \text{PSATB(T1)} \tag{B-5}$$

The pump inlet pressure is also equal to the well bottom pressure P1 minus the hydrostatic pressure difference RHOB(T1) YPUMP GRAV between the well bottom and pump inlet at height YPUMP, minus the friction pressure drop $\frac{1}{2}$ CF RHOB(T1) VL^2 π DIA YPUMP/AP, where RHOB(T1) is the brine density at well bottom temperature and CF is the skin friction coefficient. Solving for YPUMP, the height of the pump above well bottom is

$$YPUMP = \frac{P1 - PIN}{RHOB(T1)\ (GRAV + 2\ CF\ VL^2/DIA)} \qquad (B\text{-}6)$$

The pump depth is

$$DPUMP = DEPTH - YPUMP \qquad (B\text{-}7)$$

F. Pump Pressure Rise and Brine Cooling

In accordance with Assumption 6, the wellhead pressure PW is equal to the saturation pressure at the wellhead temperature TW.

$$PW = PSATB(TW) \qquad (B\text{-}8)$$

The wellhead pressure is also equal to the pump inlet pressure PIN minus the hydrostatic pressure difference RHOB(TW) DPUMP GRAV between the pump and wellhead, minus the friction pressure drop $\frac{1}{2}$ CF RHOB(TW) VL^2 π DIA PUMP/AP, plus the pump pressure rise DPPUMP. Solving for DPPUMP, the pump pressure rise is

$$DPPUMP = PW - PIN + RHOB(TW)\ DPUMP\ (GRAV + 2\ CF\ VL^2/DIA) \qquad (B\text{-}9)$$

where the velocity VL is recalculated from Eq. B-4 for the density at the wellhead temperature.

In accordance with Assumption 5, the brine is cooled in the heat exchanger above the pump by amount DTPUMP given by

$$DTPUMP = KT\ \ DPPUMP \qquad (B\text{-}10)$$

where KT is the specified brine temperature drop per unit of pump pressure rise.

G. Wellhead Conditions

To find the wellhead conditions, the wellhead temperature is first approximated by the well bottom temperature T1. The corresponding wellhead pressure is found from Eq. (B-8), the pump pressure rise from Eq. (B-9), and the brine temperature drop from Eq. (B-10). If there is heat transfer out of the well there is a further temperature drop DTWELL = HTC π DIA DEPTH (TBAR - TSINK) (B-11)

where HTC is the heat transfer coefficient, TBAR is the mean brine temperature

$$TBAR = \tfrac{1}{2}(T1 + TW) \tag{B-12}$$

and TSINK is the mean sink temperature

$$TSINK = \tfrac{1}{2}(TRES + 20) \tag{B-13}$$

The wellhead temperature is equal to the well bottom temperature minus the temperature drop for the pump and for the well heat loss.

$$TW = T1 - DTPUMP - DTWELL \tag{B-14}$$

This new value of wellhead temperature is substituted in Eq. (B-8), and the procedure from Eq. (B-9) through Eq. (B-14) is repeated until the wellhead temperature converges to the correct value.

H. Pumping Power

The pumping power PPUMP is the product of volume flow rate and pump pressure rise. The density of the liquid being pumped is RHOB(T1) since the brine is not cooled until after leaving the pump (neglecting well heat loss). Hence, the pumping power is

$$PPUMP = \frac{MW \quad DDPPUMP}{RHOB(T1)} \tag{B-15}$$

I. Available Wellhead Power

The specific available wellhead power is the enthalpy change for an isentropic expansion of saturated liquid from wellhead temperature TW to rejection temperature TR. That enthalpy change is equal to the difference in saturated liquid enthalpies between the two temperatures, minus the product of the absolute rejection temperature and the difference in saturated liquid entropies.

$$PAV = HB(TW) - HB(TR) - (TR + 273.18)(SB(TW) - SB(TR)) \quad \text{(B-16)}$$

The total available wellhead power is the product of PAV(TW) and the well flow rate MW.

$$PAVT = MW \; PAV \quad \text{(B-17)}$$

J. Computer Program

A Fortran IV computer program incorporating the equations derived in this Appendix B is presented on the following pages. Results for a test case are also given.

```
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C  * * * * * * * * * * PUMPED GEOTHERMAL WELL  * * * * * * * * * * * *
C
C  * * * * * * * * * * D. G. ELLIOTT, MAY 30, 1975 * * * * * * * * * * *
C
C  * * * * * * * * * * LAST REVISION, AUG. 7, 1975 * * * * * * * * * * *
C
C  CALCULATES THE WELLHEAD CONDITIONS FOR A PUMPED GEOTHERMAL
C  WELL GIVEN THE DEPTH, RESERVOIR TEMPERATURE AND PRESSURE,
C  BRINE CONCENTRATION, WELL DIAMETER, FLOW RATE, AND THE
C  COOLING OF THE BRINE PER UNIT OF PUMP PRESSURE RISE.
C
C  * * * * * * * * EQUATION NUMBERS REFER TO APPENDIX B  * * * * * * * *
C
C  INPUT DATA
C  ----------
C
C  CONC     BRINE CONCENTRATION FRACTION (DENVER RESEARCH INSTITUTE MIX)
C  TRES     RESERVOIR TEMPERATURE, DEG C
C  PRES     RESERVOIR PRESSURE, PA (SEE NOTE 1)
C  MW       WELL FLOW RATE, KG/S
C  DIA      WELL DIAMETER, M
C  DEPTH    WELL DEPTH, M
C  KD       DRAWDOWN PRESSURE FACTOR, PA PER KG/S
C  CF       SKIN FRICTION COEFFICIENT
C  HTC      HEAT TRANSFER COEFFICIENT, J/M*2 PER DEG C (SEE NOTE 2)
C  TR       REJECTION TEMPERATURE FOR AVAILABLE POWER EVALUATION
C  KT       BRINE TEMP REDUCTION PER UNIT PUMPING PRESSURE, DEG C PER PA
C  NPR      '0' FOR NOTE-1 RESERVOIR PRESSURE, '1' FOR INPUT
C  NINPT    '1' FOR ECHO OF INPUTS
C
C  OUTPUTS
C  -------
C
C  TW, C          WELLHEAD TEMPERATURE, DEG C
C  DPPUMP,MPA     PUMP PRESSURE RISE, MPA
C  AVAIL', MW     SPECIFIC AVAILABLE WELLHEAD POWER, J/G
C  AVAIL, MW      TOTAL AVAILABLE WELLHEAD POWER, MW
C  PPUMP, MW      PUMPING POWER (VOLUME FLOW RATE * DPPUMP), MW
C  DPUMP, M       PUMP DEPTH FOR SATURATION PRESS AT INLET
C
C  NOTE 1: IF NPR=0, RESERVOIR PRESS IS SET EQUAL TO THE NORMAL HYDRO-
C          STATIC VALUE = COLUMN OF 20 DEG C PURE WATER OF WELL DEPTH.
C
C  NOTE 2: SINK TEMP ASSUMED LINEAR FROM 20 C AT TOP TO TRES AT BOTTOM.
C
      PARAMETER MXT=300 @ MAX TEMP IN PROPERTY TABLES, CAN USE LESS
C
      INTEGER TRES,T1,TR,TLO,THI
      IMPLICIT REAL(M)
      REAL PSAT(MXT),HL(MXT),HG(MXT),SL(MXT),SG(MXT),RHOL(MXT),
     &     RHOG(MXT),PSATB(MXT),HB(MXT),HGB(MXT),SL(MXT),SGB(MXT),
     &     RHOB(MXT),RHOGB(MXT),KD,KT
C
C     EQUIVALENCE (PSAT,PSATB),(HL,HB),(HG,HGB),(SL,SB),(SG,SGB),
```

```
C     &              (RHOL,RHOB),(RHOG,RHOGB) @ FOR PURE WATER CASES
C
      NAMELIST /IN/CONC,TRES,PRES,MW,DIA,DEPTH,KD,CF,HTC,TR,KT,NPR,NINPT
C
      AVAIL(T)=HB(T)-HB(TR)-(TR+273.18)*(SB(T)-SB(TR)) @ AVAIL PWR (16)
C
C     ***** LOAD PURE WATER THERMODYNAMIC PROPERTIES *****
C
      CALL H2OPRP(PSAT, HL, HG, SL, SG, RHOL, RHOG,
     &            PSATB, HB, HGB, SB, SGB, RHOB, RHOGB, MXT)
C
C     ***** DEFAULT INPUTS AND CONSTANTS *****
C
      KD=25.0E3 @ DRAWDOWN PRESSURE OF IID NO.1 (REF. HELGESON)
      DIA=0.25 @ 25-CM CASING DIAMETER
      CF=0.008 @ SKIN FRICTION COEFFICIENT
      TR=45 @ REJECTION TEMPERATURE, DEG C
      KT=4.0E-6 @ 4 DEG C PER MPA (5 DEG F PER 100 PSI), BEST VALUE
C                 PREDICTED BY SPERRY (CALTECH CONF., 1974, P.286)
      PI=3.14159
      GRAV=9.80665 @ ACCELERATION DUE TO GRAVITY, M/S**2
C
C     ***** INPUTS *****
C
10    PRINT 15
15    FORMAT (//' ENTER')
      READ (5,IN,ERR=10,END=95) @ READ INPUTS
      IF (NPR.EQ.0) PRES=RHOL(20)*DEPTH*GRAV @ NORMAL RESERVOIR PRESS
      IF (NINPT.EQ.1) WRITE (6,IN) @ ECHO INPUTS
C
C     ***** LOAD BRINE THERMODYNAMIC PROPERTIES *****
C
      OLDCON=-1 @ TO ASSURE CALL ON FIRST PASS
      IF (CONC.NE.OLDCON) CALL BRNPRP(CONC) @ LOAD BRINE PROPERTIES
      OLDCON=CONC @ TO PREVENT REPEAT CALL IF CONC NOT CHANGED
C
C     ***** WELL BOTTOM CONDITIONS *****
C
      T1=TRES @ NO TEMP CHANGE
      P1=PRES - KD*MW @ RESERVOIR PRESS MINUS DRAWDOWN (2)
      AP=PI/4 *DIA**2 @ PIPE AREA (3)
      VL=MW/(RHOB(T1)*AP) @ LIQUID (BRINE) VELOCITY (4)
C
C     ***** PUMP DEPTH *****
C
      PIN=PSATB(T1) @ PUMP INLET PRESSURE (5)
      YPUMP=(P1-PIN)/(RHOB(T1)*(GRAV + 2*CF*VL**2/DIA)) @ PUMP HT (6)
      DPUMP=DEPTH - YPUMP @ PUMP DEPTH (7)
C
C     ***** ITERATION FOR WELLHEAD TEMPERATURE *****
C
      TW=T1 @ INITIAL APPROXIMATION TO WELLHEAD TEMPERATURE
C
            DO 30 K=1,100 @ ITERATE UP TO 100 TIMES FOR TW
            TWSV=TW @ SAVE LAST WELLHEAD TEMP
            TLO=TW @ TW TRUNCATED TO INTEGER
            THI=TLO+1 @ NEXT HIGHER INTEGER TEMP
```

```
            IF (TLO.LT.20.OR.THI.GT.MXT) GO TO 85 @ TW OUT OF RANGE
            FX=TW-TLO @ FRACTIONAL DISTANCE FROM TLO TO TW
            PW=PSATB(TLO)+FX*(PSATB(THI)-PSATB(TLO)) @ WELLHD PRESS (8)
            VL=MW/(RHOB(TLO)*AP) @LIQUID VEL ABOVE PUMP (4), PUMP DP NEXT
            DPPUMP=PW - PIN + RHOB(TLO)*DPUMP*(GRAV +2*CF*VL**2/DIA) @(9)
            IF (DPPUMP.LT.0) DPPUMP=0 @ NO PUMP NEEDED
            DTPUMP=KT*DPPUMP @ BRINE COOLING DUE TO PUMP (10)
            TBAR=(T1+TW)/2.0 @ MEAN BRINE TEMPERATURE (12)
            TSINK=(TRES+20)/2.0 @ MEAN SINK TEMP (13)
            DTWELL=HTC*PI*DIA*DEPTH*(TBAR-TSINK) @ HEAT-LOSS COOLING (11)
            TW=T1 - DTPUMP - DTWELL @ WELLHEAD TEMP (14)
30          IF (ABS(TW-TWSV).LT.0.01) GO TO 40 @ CONVERGED TO 0.01 DEG C
C
      PRINT 35
35    FORMAT (' DID NOT CONVERGE ON TW')
C
40    PPUMP=MW*DPPUMP/RHOB(T1) @ PUMPING POWER (15)
      PAV=AVAIL(TLO)+FX*(AVAIL(THI)-AVAIL(TLO)) @ AVAIL PWR (16)
      PAVT=MW*PAV @ WELLHEAD TOTAL AVAILABLE POWER (17)
C
      PRINT 50 @ HEADING
50    FORMAT (/'    TW     DPPUMP  AVAIL'' AVAIL   PPUMP   DPUMP'/
     &          '     C       MPA    J/G      MW       MW      M'/)
C
      PRINT 55,TW,DPPUMP,PAV,PAVT,PPUMP,DPUMP
55    FORMAT (F7.1,-6PF8.3,-3PF7.1,2(-6PF8.3),0PF8.1//)
C
      GO TO 10 @ ASK FOR ANOTHER CASE
C
85    PRINT 90,TW
90    FORMAT (' TW OUT OF RANGE OF PROPERTY TABLES AT',F12.3)
95    STOP
      END
```

TEST CASE

$IN CONC = 0.2, TRES = 250, MW = 200, DEPTH = 1500 $END

```
ENTER
$IN
CONC     =      .20000000E+00
TRES     =              +250
PRES     =      .14683544E+08
MW       =      .20000000E+03
DIA      =      .25000000E+00
DEPTH    =      .15000000E+04
KD       =      .25000000E+05
CF       =      .80000000E-02
HTC      =      .00000000E+00
TR       =               +45
KT       =      .40000000E-05
NPR      =                +0
NINPT    =                +1
$END
```

TW C	DPPUMP MPA	AVAIL, J/G	AVAIL MW	PPUMP MW	DPUMP M
216.4	8.389	99.9	19.985	1.738	907.2

APPENDIX C

CALCULATION METHOD FOR BRINE PROPERTIES

A. Nomenclature

CONC	salt fraction by mass
DH	enthalpy difference between superheated and saturated vapor
DP	difference between water and brine saturation pressure
DT	increase in saturation temperature due to dissolved salts
HB(T)	enthalpy of saturated brine at T
HGB(T)	enthalpy of vapor in equilibrium with saturated brine at T
HG(T)	enthalpy of vapor in equilibrium with saturated pure water at T
HL(T)	enthalpy of saturated pure water of T
PSAT(T)	saturation pressure of pure water at T
PSATB(T)	saturation pressure of brine at T
RHOB(T)	density of saturated brine at T
RHOG(T)	density of vapor in equilibrium with saturated pure water at T
RHOGB(T)	density of vapor in equilibrium with saturated brine at T
RHOL(T)	density of saturated pure water at T
SB(T)	entropy of saturated brine at T
SG(T)	entropy of vapor in equilibrium with saturated pure water at T
SGB(T)	entropy of vapor in equilibrium with saturated brine at T
SL(T)	entropy of saturated pure water at T
T	temperature
TB	saturation temperature of pure water

B. Assumptions

The assumptions used in calculating brine properties are:

1. The dissolved salts consist of KCl, $CaCl_2$, and NaCl in the ratio 1.00:1.95:3.55 by mass.
2. The saturation pressure, liquid density, and liquid enthalpy are those measured by the Denver Research Institute [C-1].
3. The vapor properties are those of superheated steam at the saturation pressure of the brine.

Assumption 1 is necessary because no other data for brine properties are available. However, the composition is typical of the geothermal brines in the Salton Sea area[C-2]. In other areas where the composition is different the concentration is usually less so that the differences between pure water and brine properties are small and the thermodynamic properties are, therefore, not greatly affected by the assumed composition.

Assumption 2 is valid for brines of the assumed composition.

Assumption 3 is accurate since the vapor from brine is nearly pure steam and the pressure is lower than the saturation pressure of pure water.

In addition to those assumptions, several numerical approximations are introduced during the derivation.

C. Pure Water Properties

The properties of saturated pure water and saturated steam are taken from the Steam Tables[C-3]. The saturation pressure, liquid enthalpy, vapor enthalpy, liquid entropy, and vapor entropy are loaded into the arrays PSAT(T), HL(T), HG(T), SL(T) and SG(T), respectively, for integer values of temperature T from 20°C to 300°C. The specific volumes of the liquid and vapor are converted to densities and loaded into the arrays RHOL(T) and RHOG(T), respectively, for

---

C-1 Pool, M.J., and Nevens, T. D., Determination of Thermodynamic Properties of Brines, Report No. 2151, Denver Research Institute, Denver, Colorado, Feb. 1964.

C-2 Helgeson, H. C., "Geologic and Thermodynamic Characteristics of the Salton Sea Geothermal System," American Journal of Science, Vol. 266, pp. 129-166, March 1968.

C-3 Keenan, J. H., and Keyes, F. G., Steam Tables (Metric Units), Wiley, New York, 1967.

the same temperatures. The arrays are in SI units: Pa for pressure, J/kg for enthalpy, J/kg^{o}C for entropy, and kg/m^{3} for density, obtained by multiplying the values in Reference C-3 by 10^{5} for pressure, 100 for enthalpy and entropy, and dividing specific volumes into 1000 to get densities.

D. Brine Properties

The brine properties are calculated from the water properties for given concentration CONC using the formulas derived in Subsections E-K. The brine saturation pressure, liquid enthalpy, vapor enthalpy, liquid entropy, vapor entropy, liquid density, and vapor density are loaded into the arrays PSAT(T), HG(T), HGB(T), SB(T), SGB(T), RHOB(T), and RHOGB(T), respectively, for integer temperatures T from 20^{o}C to 300^{o}C.

E. Saturation Pressure

Figure 6 of Reference C-1 gives the boiling point rise DT as a function of the boiling point TB of pure water and the salt concentration CONC. The data can be represented by

$$DT = 0.255\ (TB + 45)\ CONC^{1.32} \qquad (C\text{-}1)$$

where TB and DT are in ^{o}C.

Given the brine temperature T, the temperature of pure water with the same vapor pressure is TB = T - DT. From Eq. (C-1) TB is given by

$$TB = (T - 11.48\ CONC^{1.32})/(1 + 0.255\ CONC^{1.32}) \qquad (C\text{-}2)$$

The saturation pressure of brine at temperature T is the saturation pressure of pure water at temperature TB.

$$PSATB(T) = PSAT(TB) \qquad (C\text{-}3)$$

where PSAT(TB) is found by interpolating between the PSAT values at the next higher and lower integral ^{o}C.

F. Liquid Enthalpy

Figure 9 of Reference C-1 presents enthalpy data showing that the difference between brine enthalpy at 0.35 concentration and brine enthalpy at a lower concentration CONC varies approximately as $(1 - CONC/0.35)^{3}$. The difference

between brine enthalpy at 0.35 concentration and pure water enthalpy varies approximately linearly with temperature. A curve fit to Fig. 9 of Reference C-1 using those dependences is

$$HB(T) = HL(T)\ \{1 - [0.2388 + 4.275 \times 10^{-4} T][1-(1 - CONC/0.35)^3]\} \quad (C\text{-}4)$$

G. Vapor Enthalpy

The vapor in equilibrium with brine is in the superheat region because the brine saturation pressure is less than the saturation pressure of pure water at the same temperature. Making an approximate curve fit to the Steam Tables, the enthalpy increase DH relative to the saturation enthalpy HG(T) is related to the pressure drop DP below the pure water saturation pressure PSAT(T) by the formula

$$DH/HG(T) = 2.18 \times 10^{-11}\ T^4\ DP/PSAT(T) \quad (C\text{-}5)$$

with T in ^{o}C. The brine vapor enthalpy is thus

$$HGB(T) = HG(T)\ (1 + DH/HG(T)) \quad (C\text{-}6)$$

H. Liquid Density

Figure 2 of Reference C-1 gives liquid density as a function of temperature and concentration. The results can be represented by a linear density increase of 830 kg/m^3 per unit of concentration. Thus,

$$RHOB(T) = RHOL(T) + 830\ CONC \quad (C\text{-}7)$$

I. Vapor Density

The density of the superheated vapor in equilibrium with the brine can be approximated by a perfect gas expansion from PSAT(T) to PSATB(T). Thus,

$$RHOGB(T) = RHOG(T)\ [PSATB(T)/PSAT(T)] \quad (C\text{-}8)$$

J. Liquid Entropy

The entropy change of the liquid between two temperatures can be found from the enthalpy and density of the liquid and the saturation pressure using the thermodynamic relation $ds = (dh - dp/\rho)/T$. For a temperature change from T - 1 to T the liquid entropy at T is given by

$$SB(T) = SB(T-1) + \frac{HB(T) - HB(T-1) - 2\dfrac{PSATB(T) - PSATB(T-1)}{RHOB(T) + RHOB(T-1)}}{T - 0.5 + 273.18} \qquad (C\text{-}9)$$

where T - 0.5 is the mean temperature in ^{o}C and the addition of 273.18 converts the temperature to ^{o}K.

The liquid entropy at $20^{o}C$, the starting temperature for loading the properties, is found by applying Eq. (C-9) between 0 and $20^{o}C$. Since HB(0) = SB(0) = 0, PSATB (0) is negligible, RHOB(0) $\cong$ RHOB(20), and the mean temperature is $10^{o}C$, the liquid entropy at $20^{o}C$ is

$$SB(20) = \frac{HB(20) - PSATB(20)/RHOB(20)}{10 + 273.18} \qquad (C\text{-}10)$$

K. Vapor Entropy

The entropy of vaporization is equal to the enthalpy of vaporization divided by the absolute temperature. Thus, the vapor entropy is

$$SGB(T) = SB(T) + \frac{HGB(T) - HB(T)}{T + 273.18} \qquad (C\text{-}11)$$

L. Computer Program

A Fortran IV subroutine incorporating the equations derived in the Appendix is presented on the following pages. The subroutine is used by the well flow and conversion process programs and can be checked with the same test cases.

```
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C  * * SUBROUTINE FOR THERMODYNAMIC PROPERTIES OF WATER AND BRINE  * * *
C
C  * * * * * * * * D. G. ELLIOTT, APRIL 25, 1975 * * * * * * * * * * *
C
C  * * * * * * * * * * *  LAST REVISION, AUG. 8, 1975  * * * * * * * * * *
C
C  * * * * * * * * EQUATION NUMBERS REFER TO APPENDIX C  * * * * * * * *
C
C  FILE 10 HAS SATURATION PRESSURE IN BARS, LIQUID AND VAPOR ENTHALPIES
C  IN J/G, AND LIQUID AND VAPOR ENTHALPIES IN J/G*C FROM 20 C TO 300 C.
C
C  FILE 11 HAS SPECIFIC VOLUMES IN CC/G FROM 20 C TO 300 C.
C
C  SOURCE: KEENAN & KEYES 1969 STEAM TABLES (METRIC UNITS)
C
      SUBROUTINE H2OPRP(PSAT,HL,HG,SL,SG,RHOL,RHOG,PSATB,
     &                  HB,HGB,SB,SGB,RHOB,RHOGB,MXT)
C
      INTEGER T,TLO,THI
      REAL PSAT(1),HL(1),HG(1),SL(1),SG(1),RHOL(1),RHOG(1),
     &     PSATB(1),HB(1),HGB(1),SB(1),SGB(1),RHOB(1),RHOGB(1)
C
C     READ SATURATION PRESSURE, ENTHALPIES, AND ENTROPIES FROM FILE 10.
C
      READ (10,20) (TT,PSAT(T),HL(T),HG(T),SL(T),SG(T),T=20,MXT)
20    FORMAT ((F12.5,-5PF12.5,4(-3PF12.5)))
C
C     (SCALE FACTORS IN FORMAT 20 CONVERT PROPERTIES TO SI UNITS:
C     PSAT IN PA, HL AND HG IN J/KG, SL AND SG IN J/KG*C)
C
      READ (11,25) (RHOL(T),RHOG(T),T=20,MXT) @ READ VOLUMES FROM 11
25    FORMAT (2(3PF12.5))
C
C     (SCALE FACTOR IN FORMAT 25 CONVERTS SPECIFIC VOLUMES TO M**3/KG)
C
          DO 30 T=20,MXT @ CONVERT SPECIFIC VOLUMES TO DENSITIES
          RHOL(T)=1.0/RHOL(T) @ LIQUID DENSITY IN KG/M**3
30        RHOG(T)=1.0/RHOG(T) @ VAPOR DENSITY IN KG/M**3
C
      RETURN @ PURE WATER PROPERTIES LOADED
C
C  * * * * * * * * *  ENTRY FOR BRINE PROPERTIES * * * * * * * * * * * *
C
      ENTRY BRNPRP(CONC) @ LOADS PROPERTIES OF SATURATED BRINE
C     HAVING SALT MASS FRACTION = CONC.
C
C  BRINE PROPERTIES FROM POOL AND NEVENS, REPORT NO. 2151,
C  DENVER RESEARCH INSTITUTE, FEBRUARY 1964.
C
C  BRINE DENSITY LINEAR WITH CONC AT 830 KG/M**3 PER UNIT (DRI FIG. 2)
C  BRINE LIQ ENTHALPY FROM FIT TO DRI FIG. 9.
C  LIQUID ENTROPY FROM INTEGRATION OF DS=(DH-DP/RHO)/T ALONG SAT LINE
C  BRINE SAT PRESS = WATER SAT PRESS AT LOWER TEMP GIVEN BY DRI FIG. 6
C  VAPOR DENSITY FOR GAS EXPANDING FROM H2O SAT PRESS TO BRINE SAT PRESS
```

```
C   VAPOR ENTHALPY FROM FIT TO DH/DP GOING INTO SUPERHEAT REGION
C   VAPOR ENTROPY SUCH THAN ENTROPY OF VAPORIZATION =
C   (ENTHALPY OF VAPORIZATION)/(T, DEG K)
C
            DO 35 T=25,MXT @ LOAD SATURATION PRESSURES FROM 25 C
            TB=(T-11.48*CONC**1.32)/(1+0.255*CONC**1.32) @ H2O SAT T (2)
            TLO=TB @ NEXT LOWER INTEGER TEMP
            IF (TLO.LT.20) PRINT 37 @ WARN OF GOING BELOW 20 C
37          FORMAT (' WENT BELOW 20 C IN PSATB ROUTINE')
            THI=TLO+1 @ NEXT HIGHER INTEGER TEMP
35          PSATB(T)=PSAT(TLO)+(PSAT(THI)-PSAT(TLO))*(TB-TLO) @ SAT P (3)
C
            DO 40 T=20,24 @ LOAD SATURATION PRESSURES BELOW 25 C
40          PSATB(T)=PSAT(T)*PSATB(25)/PSAT(25) @ ASSUME SAME RATIO
C
      CFCTR=(1-CONC/0.35)**3 @ FCTR OF ENTHALPY DIFF FROM 0.35 CONC
      IF (CONC.GT.0.35) CFCTR=0 @ ENTHALPY CONSTANT BEYOND 0.35 CONC
C
            DO 45 T=20,MXT @ LOAD DENSITIES AND ENTHALPIES
            RHOGB(T)=RHOG(T)*PSATB(T)/PSAT(T) @ VAPOR DENSITY (8)
            RHOB(T)=RHOL(T)+830*CONC @ BRINE DENSITY (7)
            DP=PSAT(T) - PSATB(T) @ PRESS DROP BELOW H2O SATURATION
            HGB(T)=HG(T)*(1 + 2.18E-11*T**4 *DP/PSAT(T)) @CURVE FIT (5,6)
45          HB(T)=HL(T)*(1-(0.2388+4.275E-4*T)*(1-CFCTR)) @ LIQ ENTH (4)
C
      SB(20)=(HB(20)-PSATB(20)/RHOB(20))/283.18 @LIQ ENTROPY @ 20 C (10)
C
            DO 50 T=20,MXT @ LOAD ENTROPIES
            IF (T.GT.20) SB(T)=SB(T-1) +(HB(T)-HB(T-1) -2*(PSATB(T)
     &      -PSATB(T-1))/(RHOB(T)+RHOB(T-1)))/(T+272.68) @LIQ ENTROPY (9)
50          SGB(T)=SB(T)+(HGB(T)-HB(T))/(T+273.18) @ VAPOR ENTROPY (11)
C
      RETURN @ BRINE PROPERTIES LOADED
C
      END
```

APPENDIX D

CALCULATION METHOD FOR FLASH STEAM AND DUAL STEAM PROCESSES

A. Nomenclature

AVAIL	specific available wellhead power
CONC	brine concentration
DH	enthalpy change during expansion
DT	difference between inlet and exit temperatures of a stage
ETA	efficiency of an expansion process
ETAC	equivalent efficiency PNET/QIN
ETAP	efficiency of pumps to return liquid to atmospheric pressure
ETA1	efficiency of steam turbines
ETA2	efficiency of two-phase expanders
HB(T)	enthalpy of saturated brine at T
HG(T)	enthalpy of vapor in equilibrium with water at T
HGB(T)	enthalpy of vapor in equilibrium with brine at T
HL(T)	enthalpy of saturated water at T
H1	mixture enthalpy before expansion
H2	mixture enthalpy after expansion
H2I	mixture enthalpy after isentropic expansion
HG1	vapor enthalpy before expansion
HG2	vapor enthalpy after expansion
HL1	liquid enthalpy before expansion
HL2	liquid enthalpy after expansion
M	flow rate
MGFV	vapor flow rate leaving flash vaporizer or two-phase expander
MLFV	separator liquid flow plus previous-stage turbine moisture
MST	total flow rate in steam turbine
MTFV	total flow rate in flash vaporizer or two-phase expander
MG2	vapor flow rate leaving steam turbine
ML2	liquid flow rate leaving steam turbine

NSTG	number of stages
P	output power from an expansion process
PATM	atmospheric pressure, 0.1013 MPa
PC	condensing pressure
PD	saturation pressure at liquid discharge temperature
PLP	pump power to return liquid to atmospheric pressure
PNET	net output power
PRF	power recovery fraction PNET/AVAIL
PSAT(T)	saturation pressure of water at T
PSATB(T)	saturation pressure of brine at T
PSUM	sum of turbine and two-phase expander output powers
PWRX	output power of liquid exhaust expander
PWR1	steam turbine output power
PWR2	two-phase expander output power
QC	condenser heat rejection
QIN	enthalpy of wellhead flow relative to liquid at TR
RHOB(T)	density of saturated brine at T
RHOG(T)	density of vapor in equilibrium with water at T
RHOGB(T)	density of vapor in equilibrium with brine at T
RHOL(T)	density of saturated water at T
SB(T)	entropy of saturated brine at T
SG(T)	entropy of vapor in equilibrium with water at T
SGB(T)	entropy of vapor in equilibrium with brine at T
SL(T)	entropy of saturated water at T
S1	mixture entropy before expansion
SG1	vapor entropy before expansion
SG2	vapor entropy after expansion
SL1	liquid entropy before expansion
SL2	liquid entropy after expansion
TD	liquid discharge temperature
TR	steam condensing temperature
TW	wellhead temperature

TX(k)	array of stage inlet and exit temperatures
T0	flash vaporizer or two-phase expander inlet temperature
T1	flash vaporizer or two-phase expander exit temperature (steam turbine inlet temperature)
T2	steam turbine exit temperature
VOL1	volume flow rate at exit of steam turbine
VOL2	volume flow rate at exit of flash vaporizer or two-phase expander
XIN	quality at entrance of flash vaporizer or two-phase expander
XOUT	quality at exit of flash vaporizer or two-phase expander
XW	wellhead quality
XX	exit quality from optional exhaust liquid expander
X1	quality before expansion
X2	quality after expansion
X2I	quality after isentropic expansion
XS2	steam turbine exit quality

B. Process Diagrams

Figure D-1 is a diagram of one stage of a flash steam or dual steam process. There are three inlet lines. The first line carries dry, saturated steam (labeled G for gas phase) from the previous-stage turbine exhaust at temperature T1. The second line carries saturated liquid (L) from the previous-stage separator (or, in the case of the first stage, two-phase flow (L + G) from the wellhead) at a higher temperature T0. The third line carries liquid separated from the previous-stage turbine exhaust at temperature T1.

The liquid or two-phase flow in the second line enters the flash vaporizer or two-phase expander at flow rate MTFV and quality XIN. The flow expands to the lower temperature T1 and leaves the flash vaporizer or two-phase expander with vapor flow rate MGFV, quality XOUT, and volume flow rate VOL2. The efficiency of the two-phase expander is ETA2 and the power output is PWR2. For a flash vaporizer ETA2 and PWR2 are zero.

The two-phase mixture leaving the flash vaporizer or two-phase expander enters a separator. The vapor at flow rate MGFV and temperature T1 joins the flow from

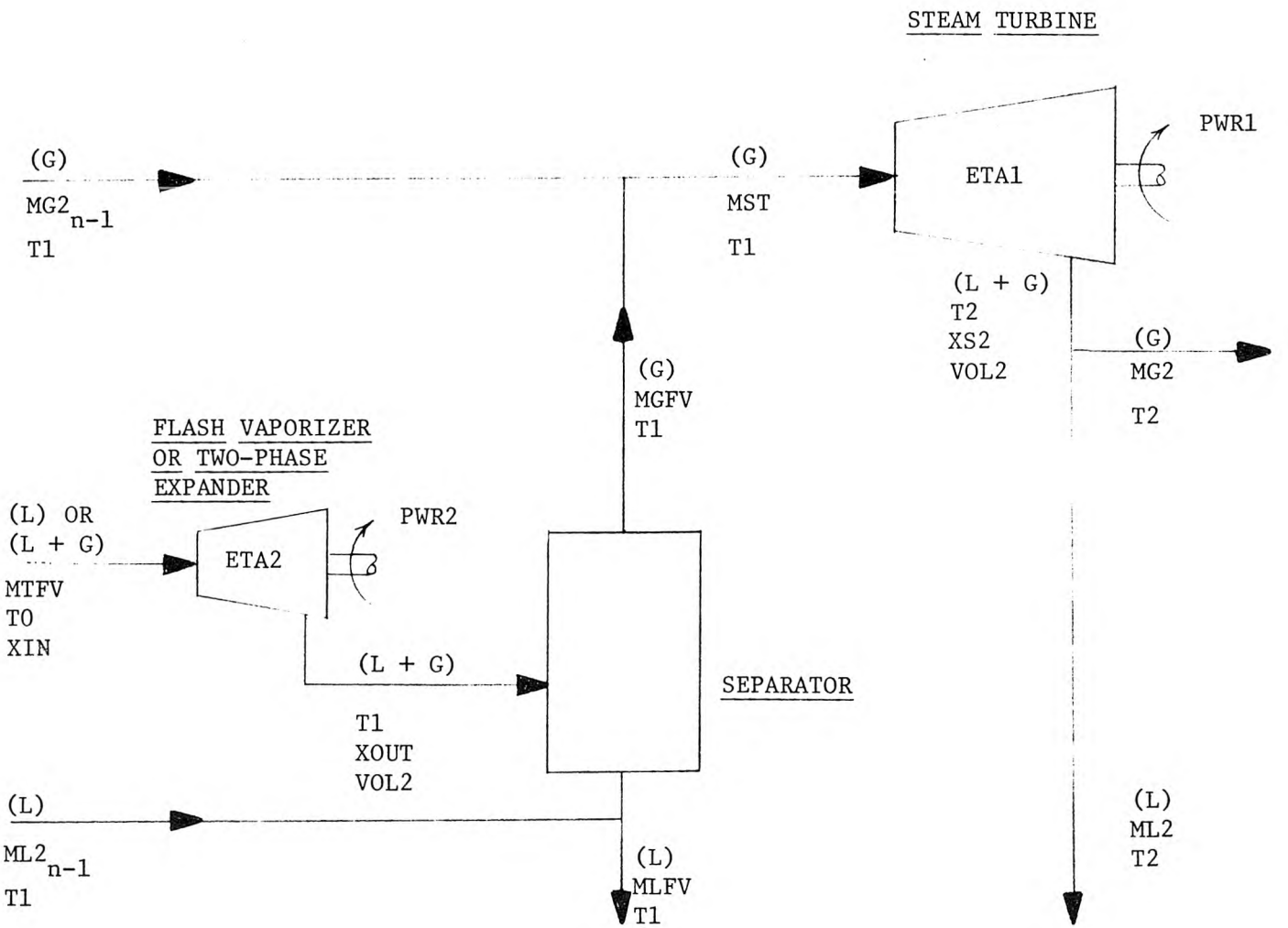

Fig. D-1 One stage of a flash steam (ETA2 = 0) or dual steam (ETA2 > 0) process

the vapor inlet line and enters the steam turbine with combined flow rate MST. The steam expands to temperature T2 and leaves the turbine with liquid flow rate ML2, vapor flow rate MG2, quality X2, and volume flow rate VOL2. The efficiency of the steam turbine is ETA1 and the output power is PWR1. The moisture at the turbine exit is separated from the vapor.

There are three discharge lines from the stage: the vapor line from the turbine with flow rate MG2 at temperature T2, the liquid line combining the separator liquid flow and previous-stage moisture flow at flow rate MLFV and temperature T1, and the moisture line from the turbine with liquid flow rate ML2 at temperature T2.

The only difference between the flash steam and dual steam processes is that ETA2 is zero for the former and greater than zero for the latter.

Figure D-2 shows how the stages are combined to form a multi-stage system. Three stages are shown; the arrangement for other numbers of stages follows the same pattern. The first stage receives the wellhead flow of temperature TW and quality XW. The flow rate, for purposes of analysis, is unity; thus, MTFV = 1 for the first stage. The brine has salt concentration CONC and the enthalpy of the flow, relative to liquid at condensing temperature TR, is QIN.

The intermediate stages are as drawn in Fig. D-1. The last stage has two possible arrangements. The liquid leaving the last-stage separator can be discarded at a temperature TD equal to the separator temperature T1, and a pump used, if necessary, to return the liquid from the corresponding saturation pressure PD to atmospheric pressure PATM. Alternatively, the liquid can be expanded to the condensing temperature TR through an optional exhaust liquid expander to provide additional output power PWRX. The vapor leaving the exhaust liquid expander is condensed with the steam leaving the steam turbine, and the liquid leaving the exhaust liquid expander is pumped to atmosphere with the condensate. The sum of the pump powers is PLP.

The temperatures at the inlet and exit of each stage are given by the array TX(1), · · ·, TX(NSTG + 2), where TX(1) = TW and TX(NSTG + 2) = TR.

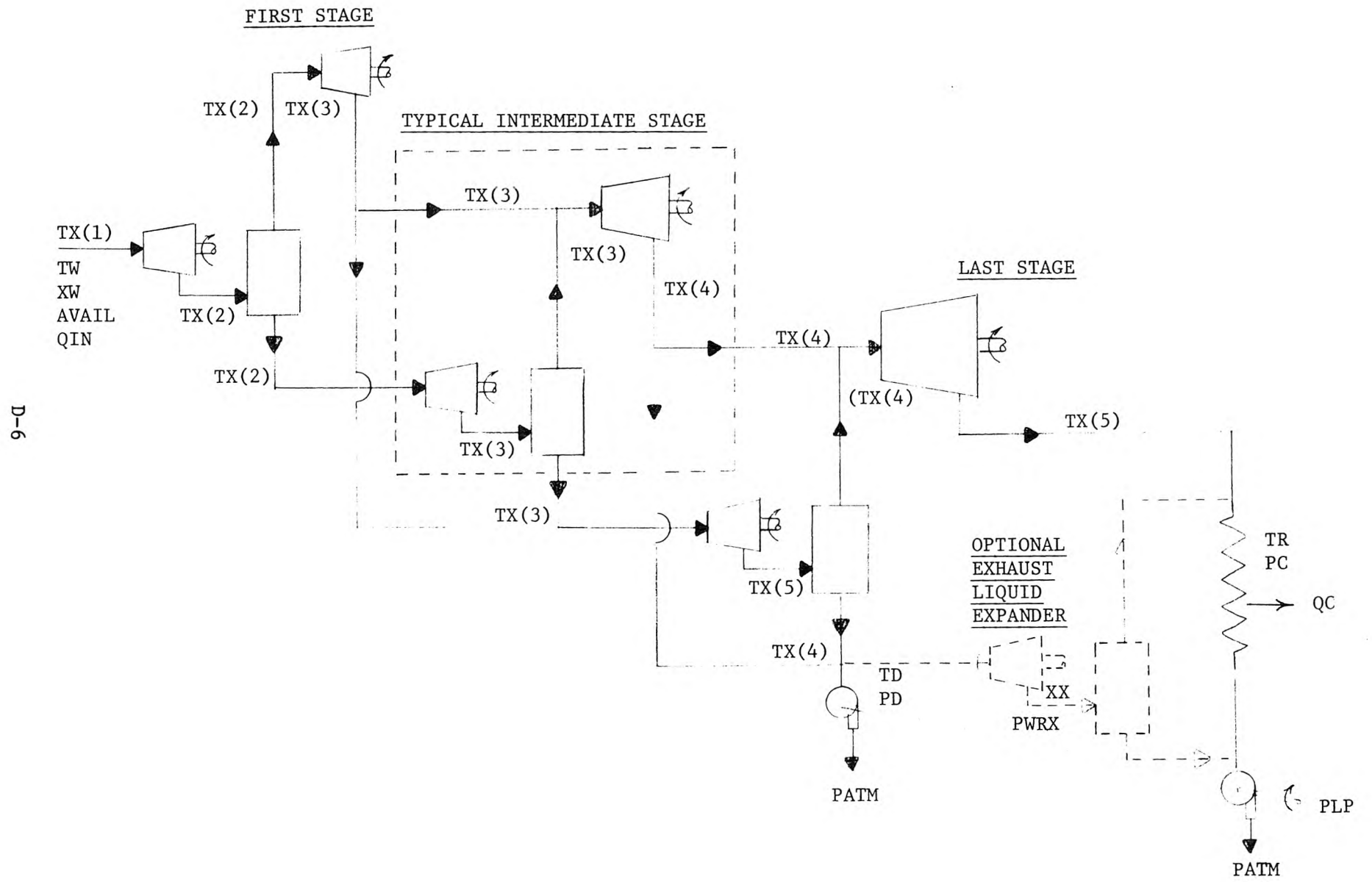

Fig. D-2 Multi-stage flash steam or dual steam process

C. Assumptions

The assumptions used in the analysis of the flash steam and dual steam processes are:

1. The inlet flow to the steam turbines is dry steam in equilibrium with brine at the same temperature.
2. The exit flow from the steam turbines is wet steam.
3. The flash vaporizer or two-phase expander exit temperature in each stage is equal to the steam turbine exit temperature for the previous stage.
4. The flow leaving the steam turbines is throttled to the lower pressure of the brine in the next stage.
5. The flow entering the flash vaporizers or two-phase expanders is saturated liquid except for the first stage in which the flow is at wellhead quality.
6. The increase in salt concentration in the liquid during the successive vaporizations is neglected.
7. There are no pressure drops except in the flash vaporizers, two-phase expanders, and steam turbines.
8. There are no heat losses except from the condenser.
9. All liquid is discharged at atmospheric pressure.
10. The steam turbines have equal efficiencies and the two-phase expanders have equal efficiencies.

Assumption 1 represents ideal operation of the separators with no carry-over of moisture or salt in the steam. Assumption 2 is valid for all practical conditions; the degree of superheat of the steam due to the reduced saturation pressure of brine is not enough to give dry steam at the turbine exits for the pressure ratios of interest.

Assumption 3 is one of two alternative choices. Since the vapor leaving the steam turbine is in equilibrium with pure water while the vapor leaving the separator in the next stage is in equilibrium with brine, it is not possible to match both temperature and pressure when the two vapor streams mix. One choice is to match pressures and mix streams of different temperatures. The choice made here is to match temperatures and have unequal pressures. The energy recoverable

from the pressure difference is small and, according to Assumption 4, is merely dissipated in throttling. The same assumptions apply to the turbine moisture stream.

Assumption 5 represents ideal operation of the separators in preventing carry-over of vapor with the liquid. Assumption 6 simplifies the analysis and introduces little error.

Assumptions 7 and 8 simplify the analysis and introduce little error since the effects of pressure drops and heat losses can be included in the turbine and two-phase expander efficiencies.

Assumption 9 meets usual pressure requirements for liquid discharge to reinjection wells, cooling towers, or surface disposal. Assumption 10 is adopted for simplicity, to allow a single efficiency value to be specified for all of the steam turbines and another for the two-phase expanders.

D. Expansion Processes

The expansion processes in the flash vaporizer, two-phase expander, or steam turbine can all be calculated by the same procedure. Given a mixture of quality X1, liquid entropy SL1, vapor entropy SG1, liquid enthalpy HL1, and vapor enthalpy HG1, the mixture entropy per unit mass is

$$S1 = (1 - X1)\ SL1 + X1\ SG1 \qquad \text{(D-1)}$$

and the mixture enthalpy per unit mass is

$$H1 = (1 - X1)\ HL1 + X1\ HG1 \qquad \text{(D-2)}$$

For an isentropic expansion to a temperature at which the liquid entropy is SL2 and the vapor entropy is SG2 the quality after expansion is

$$X2I = \frac{S1 - SL2}{SG2 - SL2} \qquad \text{(D-3)}$$

The enthalpy of the mixture per unit mass after the isentropic expansion is

$$H2I = (1 - X2I)\ HL2 + X2I\ HG2 \tag{D-4}$$

where HL2 is the liquid enthalpy and HG2 is the vapor enthalpy after expansion.

For the actual expansion at efficiency ETA the enthalpy change is

$$DH = ETA\ (H1 - H2I) \tag{D-5}$$

and the enthalpy after expansion is

$$H2 = H1 - DH \tag{D-6}$$

The corresponding quality is

$$X2 = \frac{H2 - HL2}{HG2 - HL2} \tag{D-7}$$

If the mixture flow rate is M, the output power from the expansion process is

$$P = M\ DH \tag{D-8}$$

E. Stage Analysis

The flow entering the flash vaporizer or two-phase expander (Figure D-1) has temperature T0, total flow rate MTFV, and quality XIN. The latter is equal to the wellhead quality XW for the first stage and is zero for the other stages. The mixture expands to temperature T1 at expander efficiency ETA2 (zero for a flash vaporizer). Equations (D-1)—(D-8) give the exit quality XOUT and the output power PWR2. The vapor flow rate at the exit of the flash vaporizer or two-phase expander is

$$MGFV = XOUT\ MTFV \tag{D-9}$$

and the liquid flow rate, after addition of the moisture from the previous-stage steam turbine, is

$$MLFV = MTFV - MGFV + ML2_{n-1} \tag{D-10}$$

The volume flow rate of liquid leaving the flash vaporizer or two-phase expander is (MTFV - MGFV)/RHOB(T1) and the volume flow rate of the vapor is MGFV/RHOGB(T1), where RHOB(T1) and RHOGB(T1) are the densities of the liquid and vapor, respectively, at expander exit temperature T1. The total volume flow rate is

$$VOL2 = \frac{MTFV - MGFV}{RHOB(T1)} + \frac{MGFV}{RHOGB(T1)} \tag{D-11}$$

The flow rate of vapor leaving the separator is MGFV, and the flow rate of vapor leaving the turbine of the previous stage is $MG2_{n-1}$. Thus, the flow rate entering the turbine is

$$MST = MGFV + MG2_{n-1} \tag{D-12}$$

The steam is expanded to temperature T2. Equations (D-1)—(D-8) give the exit quality XS2 and output power PWR1. The exit vapor flow is

$$MG2 = XS2\ MST \tag{D-13}$$

and the exit liquid flow is

$$ML2 = MST - MG2 \tag{D-14}$$

The sum of the liquid and vapor volume flow rates at the turbine exit is

$$VOL1 = \frac{ML2}{RHOL(T2)} + \frac{MG2}{RHOG(T2)} \tag{D-15}$$

where RHOL(T2) is the density of saturated pure water at T2 and RHOG(T2) is the density of saturated steam at T2.

If the optional exhaust liquid expander is used in the last stage, the expander inlet flow rate is MLFV, the inlet quality is zero, and the exit temperature is TR. Equations (D-1)—(D-8) give the expander exit quality XX and expander output power PWRX. The exit volume flow rate is

$$VOL2 = \frac{(1 - XX)\ MLFV}{RHOB(TR)} + \frac{XX\ MLFV}{RHOGB(TR)} \qquad (D\text{-}16)$$

F. Liquid Discharge Pumping Power

The condenser pressure PC is equal to PSAT(TR), the saturation pressure of water at the condensing temperature TR. If PC is below atmospheric pressure PATM it is necessary to pump the condensate to atmospheric pressure, and the pump input power is

$$PLP = \frac{MST\ (PATM - PC)}{RHOL(TR)\ ETAP} \qquad (D\text{-}17)$$

where RHOL(TR) is the density of water at TR and ETAP is the pump efficiency.

The brine discharge temperature TD is T1 without an exhaust expander and TR with an exhaust expander. The pressure PD is PSATB(TD), the saturation pressure at TD. If PD is below atmospheric it is necessary to pump the brine to atmospheric pressure, and the pumping power is increased to

$$PLP = PLP + \frac{MLFV\ (PATM - PD)}{RHOB(TD)\ ETAP} \qquad (D\text{-}18)$$

G. Condenser Heat Rejection

Without an exhaust expander, the flow rate of steam to the condenser is MG2 and the heat rejection is

$$QC = MG2\ [HG(TR) - HL(TR)] \qquad (D\text{-}19)$$

where HG(TR) and HL(TR) are the enthalpies of the vapor and liquid, respectively, at TR.

With an exhaust liquid expander the flow rate of steam to the condenser is increased by the vapor flow from the expander XX MLFV. The condenser heat rejection is increased to

$$QC = QC + XX\ MLFV\ [HG(TR) - HL(TR)] \tag{D-20}$$

H. Output Power, Power Recovery Fraction, and Efficiency

The total output power of the turbines and two-phase expanders, for unit wellhead flow rate, is

$$PSUM = \sum_{1}^{NSTG} PWR1 + \sum_{1}^{NSTG} PWR2 \tag{D-21}$$

Deducting pump power, the net output is

$$PNET = PSUM - PLP \tag{D-22}$$

The available work AVAIL from unit mass of the wellhead mixture (the specific available wellhead power) is given by Eqs. (D-1)—(D-8) for X1 = XW, T1 = TW, T2 = TR, ETA = 1, and M = 1. The power recovery fraction is the ratio of net output power to wellhead available power

$$PRF = \frac{PNET}{AVAIL} \tag{D-23}$$

The quantity that can be considered the equivalent heat input to the process is the difference between the enthalpy of the wellhead mixture and the enthalpy of brine at the condensing temperature TR, since that is the heat that would have to be supplied to heat the brine from TR to the wellhead temperature. That heat is

$$QIN = (1 - XW)\ HB(TW) + XW\ HGB(TW) - HB(TR) \tag{D-24}$$

The corresponding cycle efficiency of the process is

$$ETAC = \frac{PNET}{QIN} \tag{D-25}$$

I. Optimization of Temperatures

The stage inlet and exit temperatures are stored in the array TX(1), TX(2), · · ·, TX(NSTG + 2). The first temperature is the wellhead temperature

$$TX(1) = TW \tag{D-26}$$

and the last temperature is the condensing temperature

$$TX(NSTG + 2) = TR \tag{D-27}$$

If the intermediate temperatures are specified, the procedure derived in Subsections D—H gives the power recovery fraction PRF and other performance parameters of the process. Alternatively, a search procedure can be used to find the optimum values of TX(2) through TX(NSTG + 1) corresponding to maximum power recovery fraction. As an initial guess, equal temperature drops per stage are assumed.

$$DT = \frac{TW - TR}{NSTG + 1} \tag{D-28}$$

and the TX array is set to TW, TW - DT, TW - 2DT, · · ·, TR. The values of TX(2) through TX(NSTG + 1) are then varied to find the combination of temperatures that gives the maximum power recovery fraction PRF. A standard multivariate search routine is used.[D-1]

D-1 Zangwill, W., "Minimizing a Function Without Calculating Derivatives", Computer Journal, Vol. 10, 1967, pp. 293-296.

J. Computer Program

A Fortran IV computer program incorporating the equations derived in the Appendix is presented on the following pages. Results for a test case are also given.

```
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C  * * * * * * * *  FLASH STEAM AND DUAL STEAM PROCESSES * * * * * * * *
C
C  * * * * * * * * * * D. G. ELLIOTT, MARCH 26, 1975 * * * * * * * * * *
C
C  * * * * * * * * * *  LAST REVISION AUG. 12, 1975  * * * * * * * * * *
C
C  CALCULATES THE POWER GENERATED BY STEAM TURBINES AND, OPTIONALLY,
C  TWO-PHASE EXPANDERS IN A FLASHED STEAM PROCESS OF NSTG STAGES WITH
C  1 KG/S TOTAL BRINE FLOW AT WELLHEAD TEMPERATURE TW AND QUALITY XW.
C
C  * * * * * * * * EQUATION NUMBERS REFER TO APPENDIX D  * * * * * * * *
C
C  INPUT DATA
C  ----------
C
C  CONC     BRINE CONCENTRATION FRACTION (DENVER RESEARCH INSTITUTE MIX)
C  TW       WELLHEAD TEMPERATURE, DEG C
C  XW       WELLHEAD QUALITY
C  NSTG     NUMBER OF STAGES
C  TR       STEAM CONDENSING TEMPERATURE, DEG C
C  ETA1     EFFICIENCY OF STEAM TURBINES
C  ETA2     EFFICIENCY OF TWO-PHASE EXPANDERS (ZERO FOR FLASH VAPORIZERS)
C  ETAP     EFFICIENCY OF PUMPS TO RETURN LIQUID TO ATMOSPHERIC PRESSURE
C  NINPT    '1' FOR ECHO OF INPUTS
C  NSRCH    '1' FOR OPTIMIZATION OF TEMPERATURES
C  NLX      '1' FOR LIQUID EXHAUST EXPANDER IN LAST STAGE
C  TX       ARRAY OF TEMPERATURES FROM TX(1)=TW TO TX(NSTG+2)=TR
C           (SUPPLY ONLY IF NSRCH=0)
C
C  OUTPUTS (FOR 1 KG/S WELLHEAD FLOW RATE)
C  ---------------------------------------
C
C  STAGE OUTPUTS:
C
C  STG      STAGE NUMBER
C  XOUT     QUALITY AT EXIT OF FLASH VAPORIZER OR TWO-PHASE EXPANDER
C  MST      STEAM TURBINE FLOW RATE, KG/S
C  T0       TEMP AT INLET OF FLASH VAPORIZER OR TWO-PHASE EXPANDER, DEG C
C  T1       TEMP AT EXIT OF FLASH VAPORIZER OR TWO-PHASE EXPANDER, DEG C
C  T2       STEAM TURBINE EXIT TEMPERATURE, DEG C
C  XS2      STEAM TURBINE EXIT QUALITY
C  PWR1     STEAM TURBINE OUTPUT POWER, KW
C  PWR2     TWO-PHASE EXPANDER OUTPUT POWER, KW
C  VOL1     VOLUME FLOW RATE AT EXIT OF STEAM TURBINE, M**3/S
C  VOL2     VOL FLOW RATE AT EXIT OF FLASH VAP OR TWO-PHASE EXP, M**3/S
C
C  TOTALS:
C
C  PLP       PUMP POWER TO RETURN LIQUID TO ATMOSPHERIC PRESS, KW
```

```
C  PNET      NET OUTPUT POWER, (SUMMATION OF PWR1+PWR2) -PLP
C  QC        CONDENSER HEAT REJECTION, KW
C  QIN       ENTHALPY OF WELLHEAD FLOW RELATIVE TO LIQUID AT TR, KW
C  AVAIL     AVAILABLE WELLHEAD POWER, KW
C  ETAC      EQUIVALENT CYCLE EFFICIENCY, PNET/QIN
C  P/AVAIL   POWER RECOVERY FRACTION, PNET/AVAIL
C
      PARAMETER MXT=300, MXSTG=6, MXTX=MXSTG+2, N3=MXSTG*(MXSTG+3)
C
      INTEGER T0,T1,T2,TW,TR,TT,TD,TX(MXTX)
      IMPLICIT REAL (M)
      REAL PSAT(MXT),HL(MXT),HG(MXT),SL(MXT),SG(MXT),RHOL(MXT),
     &     RHOG(MXT),PSATB(MXT),HB(MXT),HGB(MXT),SB(MXT),SGB(MXT),
     &     RHOB(MXT),RHOGB(MXT),X(MXSTG),EPSX(MXSTG),WA(N3),XBEST(MXSTG)
C
C     EQUIVALENCE (PSAT,PSATB),(HL,HB),(HG,HGB),(SL,SB),(SG,SGB),
C    &            (RHOL,RHOB),(RHOG,RHOGB) @ FOR PURE WATER CASES
C
      NAMELIST /IN/CONC,TW,XW,NSTG,TR,ETA1,ETA2,ETAP,NINPT,NSRCH,NLX,TX
C
C  ***** PURE WATER THERMODYNAMIC PROPERTIES *****
C
      CALL H2OPRP(PSAT, HL, HG, SL, SG, RHOL, RHOG,
     &            PSATB, HB, HGB, SB, SGB, RHOB, RHOGB, MXT)
C
C  ***** DEFAULT INPUTS *****
C
      NSTG=1 @ ONE STAGE
      TR=45 @ 45 DEG C STEAM CONDENSING TEMPERATURE
      ETA1=0.7 @ STEAM TURBINE EFFICIENCY
      ETAP=0.7 @ LIQUID PUMP EFFICIENCY
C
C  ***** INPUTS AND CONSTANTS *****
C
10    PRINT 15
15    FORMAT (//' ENTER')
      READ (5,IN,ERR=10,END=95) @ READ INPUTS
      IF (TW.GT.MXT) PRINT 20
20    FORMAT (' MXT TOO SMALL')
      IF (NSTG.GT.MXSTG) PRINT 21
21    FORMAT (' MXSTG TOO SMALL')
      IF (TW.GT.MXT.OR.NSTG.GT.MXSTG) GO TO 10
      TX(1)=TW @ SET FIRST TEMP IN ARRAY TO WELLHEAD TEMP (26)
      TX(NSTG+2)=TR @ SET LAST TEMP TO CONDENSING TEMP (27)
      IF (NINPT.EQ.1) WRITE (6,IN) @ ECHO INPUTS
      PATM=1.013E5 @ ATMOSPHERIC PRESSURE, PA
C
C  ***** BRINE THERMODYNAMIC PROPERTIES *****
C
      OLDCON=-1 @ TO ASSURE CALL ON FIRST PASS
      IF (CONC.NE.OLDCON) CALL BRNPRP(CONC) @ LOAD BRINE PROPERTIES
      OLDCON=CONC @ TO PREVENT REPEAT CALL IF CONC NOT CHANGED
```

```
C
C  ***** AVAILABLE WELLHEAD POWER *****
C
      CALL EXPAND (1.0, XW, SB(TW), SGB(TW), SB(TR), SGB(TR), HB(TW),
     &  HGB(TW), HB(TR), HGB(TR), 1.0, XR, AVAIL) @ AVAIL PWR (1-8)
      IF (NSRCH.EQ.0) GO TO 45 @ NO OPTIMIZATION
C
C  ***** OPTIONAL OPTIMIZATION OF STAGE TEMPERATURES *****
C
      DT=(TW-TR)/(NSTG+1.0) @ GUESS AT TEMP DIFFERENCE PER STAGE (28)
C
            DO 25 K=1,NSTG @ LOAD STAGE TEMPERATURE GUESSES
25          TX(K+1)=TW-DT*K @ EQUAL TEMPERATURE STEPS
C
            DO 30 K=1,NSTG
            X(K)=TX(K+1) @ LOAD X-VECTOR WITH STAGE TEMPERATURE GUESSES
30          EPSX(K)=100.0 @ LARGE FOR NO LIMITATION ON X CHANGES
C
      EPSF=1E-3 @ CONVERGANCE CRITERION FOR POWER RECOVERY FRACTION
      ITMAX=50 @ MAX NUMBER OF SEARCH ITERATIONS
      NX=1 @ SET FLAG TO SUPPRESS PRINTOUT DURING SEARCH
      KGO=0 @ INITIALIZE SEARCH STATUS FLAG
C
35    CALL ZPMIN(NSTG,X,Y,EPSX,EPSF,ITMAX,WA,KGO,XBEST,YBEST) @ REF. D-1
      IF (KGO.GT.1) NX=0 @ REMOVE SUPPRESS FLAG FOR LAST TIME THROUGH
C
            DO 40 K=1,NSTG @ LOAD NEW STAGE TEMPERATURES FROM X VECTOR
            TT=X(K)+0.5 @ NEW INTEGER TEMPERATURE = ROUNDED X VALUE
            IF (TT.GT.TW) TT=TW @ CAN'T BE ABOVE WELLHEAD TEMPERATURE
            IF (TT.LT.TR) TT=TR @ CAN'T BE BELOW CONDENSING TEMPERATURE
40          TX(K+1)=TT @ LOAD NEW STAGE TEMPERATURE VALUES
C
C  * * * * * * *   S T A G E   C A L C U L A T I O N S   * * * * * * *
C
45    MLFV=1.0 @ WILL BE FLASH VAP OR 2-PH EXP INLET FLOW TO FIRST STAGE
      MG2=0.0 @ NO PREVIOUS VAPOR FLOW TO FIRST STAGE TURBINE
      ML2=0.0 @ NO PREVIOUS-STAGE TURBINE MOISTURE
      PSUM=0.0 @ INITIALIZE TOTAL POWER SUM TO ZERO
C
            DO 55 K=1,NSTG @ CALCULATE PERFORMANCE OF EACH STAGE
            T0=TX(K) @INLET TEMP TO FLASH VAPORIZER OR TWO-PHASE EXPANDER
            T1=TX(K+1) @ EXIT TEMP OF FLASH VAP OR TWO-PHASE EXPANDER
            T2=TX(K+2) @ STEAM TURBINE EXIT TEMP
C
C  ***** FLASH VAPORIZER (ETA2=0) OR TWO-PHASE EXPANDER (ETA2>0) *****
C
            XIN=0 @ QUALITY FOR ALL-LIQUID INLET
            IF (K.EQ.1) XIN=XW @ USE WELLHEAD QUALITY FOR FIRST STAGE
            MTFV=MLFV @ TOTAL INLET FLOW =PREV STAGE LIQUID OUT FLOW
            CALL EXPAND (MTFV, XIN, SB(T0), SGB(T0), SB(T1), SGB(T1),
     &      HB(T0), HGB(T0), HB(T1), HGB(T1), ETA2, XOUT, PWR2) @ (1-8)
            MGFV=XOUT*MTFV @ VAPOR FLOW LEAVING FLASH VAPORIZER (9)
```

```
            MLFV=MTFV-MGFV+ML2 @ SEP LIQUID FLOW PLUS PREV-STAGE MOISTURE
            VOL2=(MTFV-MGFV)/RHOB(T1)+MGFV/RHOGB(T1) @ EXIT VOL FLOW (11)
C
C   ***** OPTIONAL LAST-STAGE EXHAUST LIQUID EXPANDER *****
C
            IF (NLX.EQ.1.AND.K.EQ.NSTG)
     &      CALL EXPAND (MLFV, 0.0, SB(T1), SGB(T1), SB(TR), SGB(TR),
     &      HB(T1), HGB(T1), HB(TR), HGB(TR), ETA2, XX, PWRX) @ (1-8)
            IF (NLX.EQ.1.AND.K.EQ.NSTG) PWR2=PWR2+PWRX @ ADDED PWR
            IF (NLX.EQ.1.AND.K.EQ.NSTG)
     &      VOL2=(1-XX)*MLFV/RHOB(TR) + XX*MLFV/RHOGB(TR) @ EXIT VOL (16)
C
C   ***** STEAM TURBINE *****
C
            MST=MGFV + MG2 @ TOTAL FLOW RATE IN STEAM TURBINE (12)
            CALL EXPAND (MST, 1.0, SB(T1), SGB(T1), SL(T2), SG(T2),
     &      HB(T1), HGB(T1), HL(T2), HG(T2), ETA1, XS2, PWR1) @ (1-8)
            MG2=XS2*MST @ VAPOR FLOW AT TURBINE EXIT (13)
            ML2=MST-MG2 @ LIQUID FLOW AT TURBINE EXIT (14)
            VOL1=ML2/RHOL(T2) + MG2/RHOG(T2) @ EXIT VOL FLOW (15)
            PSUM=PSUM+PWR1+PWR2 @ CUMULATIVE TURB AND EXPANDER POWER (21)
C
C   ***** PRINTOUT OF STAGE CONDITIONS *****
C
            IF (K.EQ.1.AND.NX.EQ.0) PRINT 50 @ HEADING
50          FORMAT (//
     &      ' STG  XOUT    MST    T0    T1    T2     XS2    PWR1    PWR2',
     &      '     VOL1     VOL2'/)
55          IF (NX.EQ.0) PRINT 60,K,XOUT,MST,T0,T1,T2,XS2,PWR1,PWR2,
     &      VOL1,VOL2
60     FORMAT (I3, F7.3, F6.3, 3(I5), F7.3, 2(-3PF7.1),2(0PF8.3))
C
C   ***** PUMP POWER *****
C
       PC=PSAT(TR) @ CONDENSING PRESSURE
       PLP=0.0 @ PUMP PWR IF PC AND PD BOTH ABOVE ATMOSPHERIC
       IF (PC.LT.PATM) PLP=MST*(PATM-PC)/(RHOL(TR)*ETAP) @ COND PUMP (17)
       IF (NLX.EQ.0) TD=T1 @ LIQ DISCHARGE TEMP W/O EXPANDER
       IF (NLX.EQ.1) TD=TR @ LIQ DISCHARGE TEMP WITH EXPANDER
       PD=PSATB(TD) @ LIQUID DISCHARGE PRESSURE
       IF (PD.LT.PATM) PLP=PLP+MLFV*(PATM-PD)/(RHOB(TD)*ETAP) @LIQ P (18)
C
C   ***** POWER RECOVERY FRACTION *****
C
       PNET=PSUM-PLP @ NET OUTPUT POWER (22)
       PRF=PNET/AVAIL @ POWER RECOVERY FRACTION (23)
       NP1=NSTG+1
       IF (NSRCH.EQ.3.AND.NX.GT.0) PRINT 65,(TX(K),K=2,NP1) @ MONITOR T'S
65     FORMAT (10(I4))
       IF (NSRCH.GE.2.AND.NX.GT.0) PRINT 75,PRF @ MONITOR PRF
75     FORMAT (F7.4)
       Y=1.0-PRF @ QUANTITY BEING MINIMIZED BY ZPMIN
```

```
      IF (NX.GT.0) GO TO 35 @ RETURN TO SEARCH ROUTINE
C
C   ***** CONDENSER HEAT REJECTION AND EQUIVALENT CYCLE EFFICIENCY *****
C
      QC=MG2*(HG(TR)-HL(TR)) @ CONDENSER HEAT REJECTION (19)
      IF (NLX.EQ.1) QC=QC+XX*MLFV*(HG(TR)-HL(TR)) @WITH EXHAUST EXP (20)
      QIN=(1-XW)*HB(TW) + XW*HGB(TW) - HB(TR) @ EQUIV HEAT INPUT (24)
      ETAC=PNET/QIN @ EQUIVALENT CYCLE EFFICIENCY (25)
C
C   ***** PRINTOUT OF TOTALS *****
C
      PRINT 80 @ HEADING
80    FORMAT (//
     & ' PLP     PNET      QC       QIN      AVAIL   ETAC  P/AVAIL'/)
      PRINT 85,PLP,PNET,QC,QIN,AVAIL,ETAC,PRF @ PRINT TOTALS
85    FORMAT(-3PF5.2, 4(F9.2), 2(0PF7.4))
      GO TO 10 @ ASK FOR NEW INPUTS
95    STOP
C
C   * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C   SUBROUTINE EXPAND CALCULATES THE EXIT QUALITY AND POWER OUTPUT OF A
C   TWO PHASE EXPANSION PROCESS GIVEN THE FLOW RATE, INLET QUALITY,
C   INLET AND EXIT THERMODYNAMIC PROPERTIES, AND EFFICIENCY.
C
      SUBROUTINE EXPAND (M, X1, SL1, SG1, SL2, SG2,
     &                   HL1, HG1, HL2, HG2, ETA, X2, P)
C
      S1=(1-X1)*SL1 + X1*SG1 @ INLET MIXTURE ENTROPY (1)
      H1=(1-X1)*HL1 + X1*HG1 @ INLET MIXTURE ENTHALPY (2)
      X2I=(S1-SL2)/(SG2-SL2) @ ISENTROPIC EXIT QUALITY (3)
      H2I=(1-X2I)*HL2 + X2I*HG2 @ ISENTROPIC EXIT ENTHALPY (4)
      DH=ETA*(H1-H2I) @ ACTUAL ENTHALPY CHANGE (5)
      H2=H1-DH @ EXIT ENTHALPY (6)
      X2=(H2-HL2)/(HG2-HL2) @ EXIT QUALITY (7)
      P=M*DH @ EXPANDER POWER OUTPUT (8)
      RETURN
C
      END
```

```
$IN CONC=0.1, TW=210, WX=0.05, NSTG=3, ETA2=0.6, NINPT=1,
TX=210, 163, 113, 60, 45 $END
```

```
ENTER
$IN
CONC    =     .10000000E+00
TW      =              +210
XW      =     .50000000E-01
NSTG    =               +3
TR      =              +45
ETA1    =     .70000000E+00
ETA2    =     .60000000E+00
ETAP    =     .70000000E+00
NINPT   =               +1
NSRCH   =               +0
NLX     =               +0
TX      =             +210.          +163.          +113.          +60
                       +45.            +0.            +0.           +0
$END
```

STG	XOUT	MST	T0	T1	T2	XS2	PWR1	PWR2	VOL1	VOL2
1	.113	.113	210	163	113	.943	22.0	11.5	.117	.035
2	.069	.168	163	113	60	.938	39.3	5.5	1.209	.073
3	.071	.217	113	60	45	.978	16.9	6.5	3.236	.482

PLP	PNET	QC	QIN	AVAIL	ETAC	P/AVAIL
.12	101.64	507.87	657.50	144.09	.1546	.7054

APPENDIX E

CALCULATION METHOD FOR BINARY
AND FLASH BINARY PROCESSES

A. Nomenclature

AVAIL	specific available wellhead power
CONC	brine concentration
DT	difference between inlet and exit temperature of heat exchanger
ETAB	ratio of organic cycle output power to available power in the organic working fluid
ETAC	equivalent efficiency PNET/PIN
ETAP	efficiency of pumps to return liquid to atmospheric pressure
HB(T)	enthalpy of saturated brine at T
HF(T)	enthalpy of organic fluid at temperature T and heat exchanger pressure
HF1	enthalpy of organic fluid at entrance of heat exchanger
HF2	enthalpy of organic fluid at exit of heat exchanger
HGB(T)	enthalpy of vapor in equilibrium with brine at T
HL(T)	enthalpy of saturated water at T
MF	organic fluid flow rate
MGFV	vapor flow rate leaving flash vaporizer
MHX	heat exchanger water-side flow rate
MLFV	liquid flow rate leaving separator
MTFV	total flow rate in flash vaporizer
NSTG	number of stages
PATM	atmospheric pressure, 0.1013 MPa
PB	available power in the organic fluid
PD	saturation pressure of liquid at discharge temperature TD
PLP	pump power to return liquid to atmospheric pressure
PNET	net output power
PRF	power recovery friction PNET/AVAIL
PSATB(T)	saturation pressure of brine at T
QC	condenser heat rejection
QGX	heat transfer rate in vapor heat exchanger
QHX	heat transfer rate in one stage
QIN	enthalpy of wellhead flow relative to liquid at TR
QLX	heat transfer rate in optional last-stage liquid heat exchanger

QSUM	sum of heat exchanger heat transfer rates
RHOB(T)	density of saturated brine at T
RHOL(T)	density of saturated water at T
SF(T)	entropy of organic fluid at temperature T and heat exchanger pressure
TD	discharge temperature of brine from last-stage heat exchanger
TF(H)	temperature of organic fluid at entholpy H and heat exchanger pressure
TFMAX	maximum temperature of the organic fluid
TF1	heat exchanger inlet temperature on organic side
TF2	heat exchanger exit temperature on organic side
TPINCH	pinch-point temperature difference bewteen the water side and organic side of the heat exchangers
TR	organic fluid condensing temperature
TRA	absolute rejection temperature
TW	wellhead temperature
TX(k)	array of stage inlet and exit temperatures
T0	flash vaporizer inlet temperature
T1	heat exchanger inlet temperature on water side (flash vaporizer exit temperature)
T2	heat exchanger exit temperature on water side
XHX	quality at inlet of heat exchanger
XIN	quality at entrance of flash vaporizer
XOUT	quality at exit of flash vaporizer
XW	wellhead quality

B. Process Diagrams

Figure E-1 is a diagram of one stage of a binary or flash binary process. There are two inlet lines. The first line carries subcooled liquid (L) from the previous-stage heat exchanger exit at temperature T1. The second line carries saturated liquid (L) from the previous-stage separator (or, in the case of the first stage, two-phase flow (L+G) from the wellhead) at a higher temperature T0.

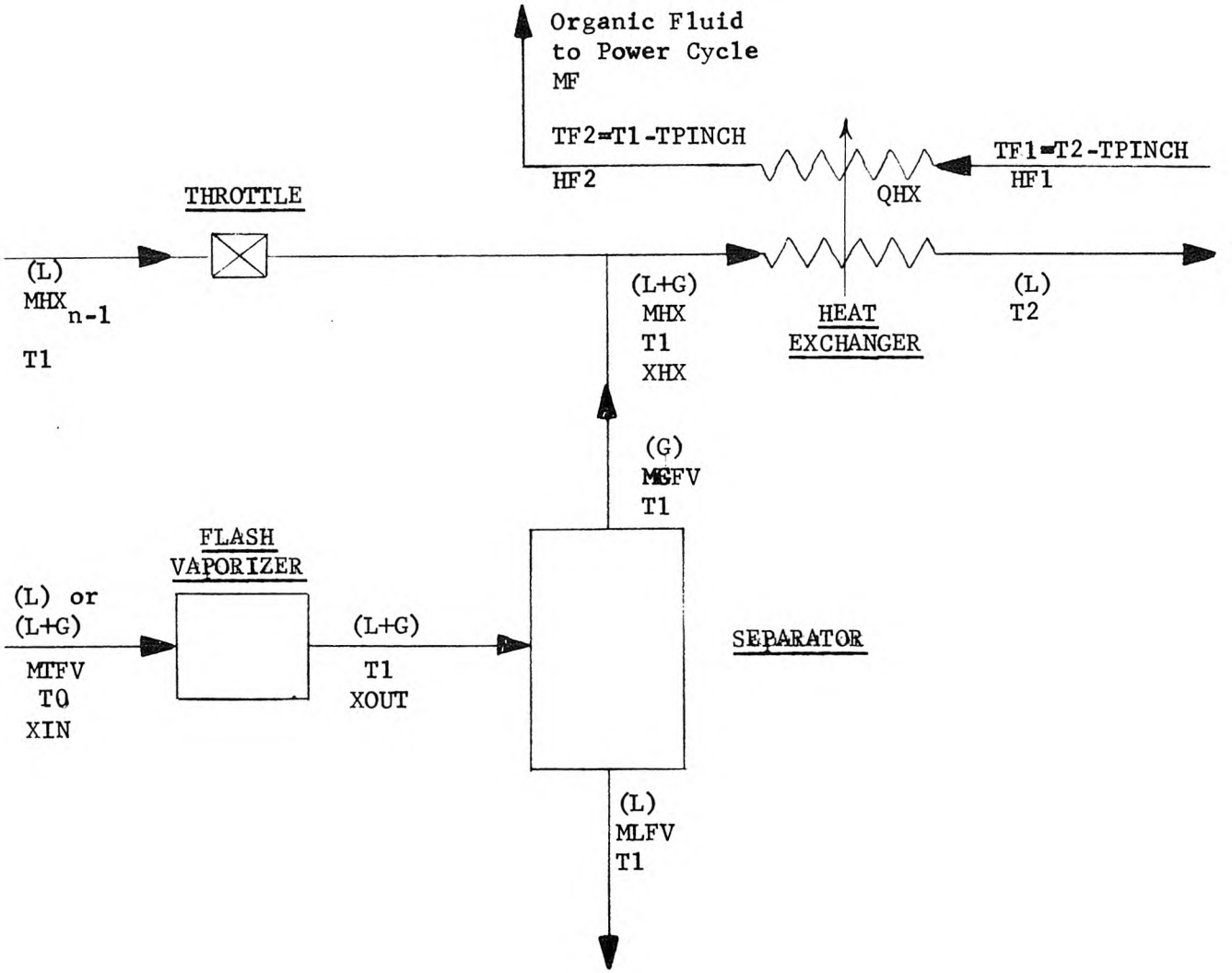

Fig. E-1 One stage of a binary or flash binary process

The liquid or two-phase flow in the second line enters the flash vaporizer at flow rate MTFV and quality XIN. The flow expands to the lower temperature T1 and leaves the flash vaporizer with vapor flow rate MGFV, liquid flow rate MLFV, and quality XOUT.

The two-phase mixture leaving the flash vaporizer enters a separator. The vapor at flow rate MGFV and temperature T1 joins the flow from the inlet line, after that flow has been throttled to the same pressure, and enters the heat exchanger with combined flow rate MHX and quality XHX. The vapor is condensed and the liquid is cooled by heat transfer to the power-cycle working fluid. The liquid leaves the heat exchanger at temperature T2 and at the saturation pressure corresponding to T1.

There are two discharge lines from the stage: the liquid line from the heat exchanger with flow rate MHX at temperature T2, and the liquid line from the separator with flow rate MLFV and temperature T1.

The power-cycle working fluid is water or an organic fluid; the term "organic fluid" will be used here. The organic fluid flows counterflow to the water, entering at temperature TF1 and leaving at TF2. The corresponding organic fluid enthalpies are HF1 and HF2. The pinch points (the locations where the water and organic fluid temperatures are closest together) are assumed to be at the inlet and exit of the heat exchanger, and the temperature difference is TPINCH. The heat transfer rate between the water and organic fluid in the stage is QHX.

Figure E-2 shows how the stages are combined to form a multi-stage system. Three stages are shown; the arrangement for other numbers of stages follows the same pattern. The first stage receives the wellhead flow of temperature TW and quality XW. The flow rate, for purposes of analysis, is unity; thus MTFV=1 for the first stage. The brine has salt concentration CONC and the enthalpy of the flow, relative to liquid at condensing temperature TR, is QIN.

The intermediate stages are as drawn in Fig. E-1. The last stage has two possible arrangements. The liquid leaving the last-stage separator can be

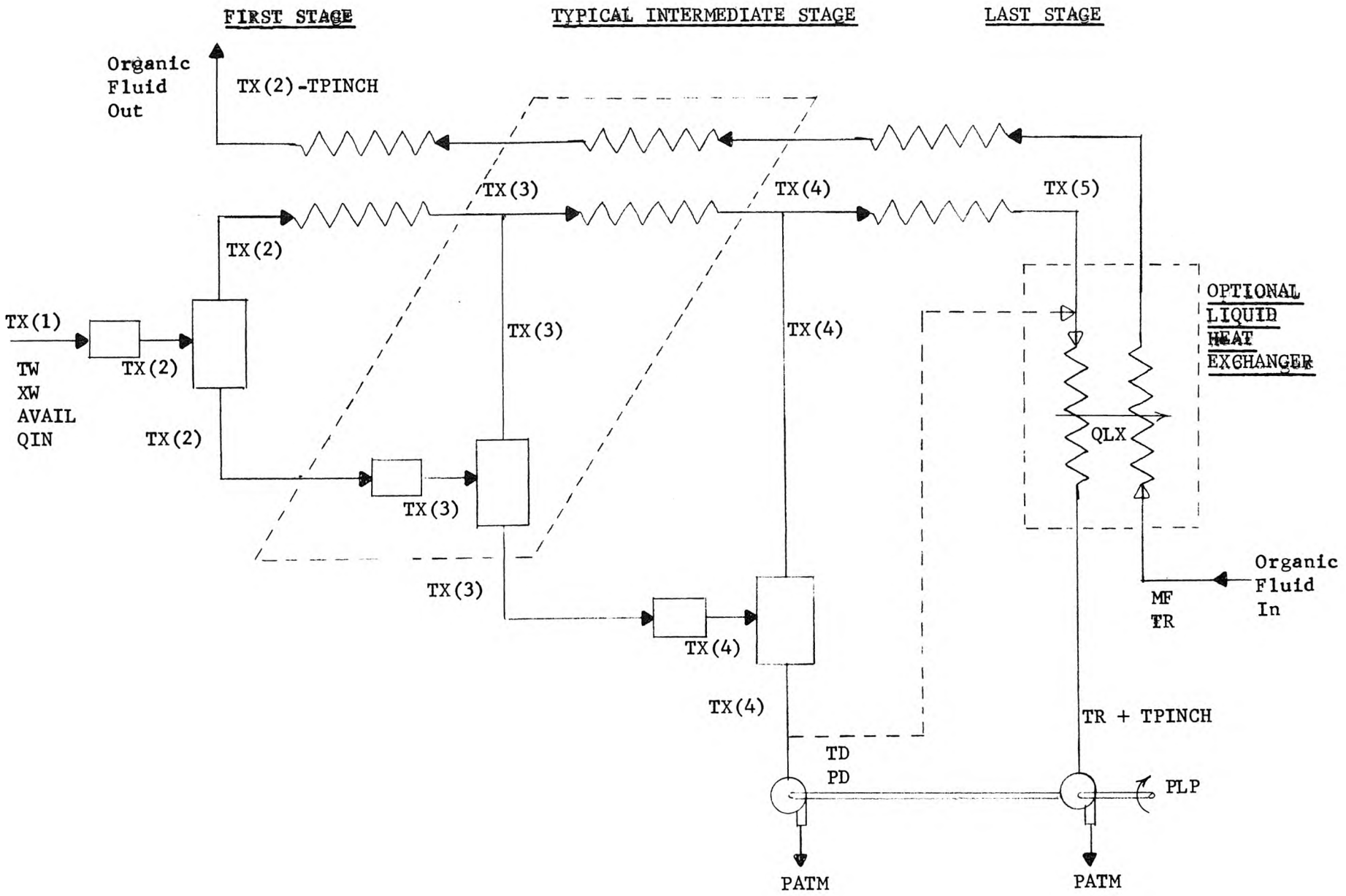

Fig. E-2 Multi-stage binary or flash binary process

discarded at a temperature TD equal to the separator temperature T1 and a pump used, if necessary, to return the liquid from the corresponding saturation pressure PD to atmospheric pressure PATM. Alternatively, the liquid can be cooled to the organic fluid condensing temperature TR plus heat-exchanger temperature difference TPINCH to provide additional heat transfer QLX. The liquid is pumped to atmosphere with the condensate from the previous heat exchanger. The sum of the pump powers is PLP.

The number of stages is NSTG. The temperatures at the inlet and exit of each stage are given by the array TX(1),..., TX(NSTG+2), where TX(1) = TW and TX(NSTG+2) = TR + TPINCH. The maximum temperature of the organic fluid is the first-stage separator temperature TX(2) minus the pinch point temperature difference TPINCH.

The only difference between the binary and flash binary processes is that the binary process has the liquid heat exchanger in the last stage.

C. Assumptions

The assumptions used in the analysis of the binary and flash binary processes are:

1. The flow to the heat exchangers is dry steam in equilibrium with brine (except for the optional last-stage liquid heat exchanger).
2. The exit flow from the heat exchangers is water (except for the last-stage liquid heat exchanger).
3. The flash vaporizer exit temperature in each stage is equal to the heat exchanger exit temperature for the previous stage.
4. The flow leaving the heat exchangers is throttled to the lower pressure of the brine in the next stage.
5. The flow entering the flash vaporizers is saturated liquid except for the first stage in which the flow is at wellhead quality.
6. The increase in salt concentration in the liquid during the successive vaporizations is neglected.
7. There are no pressure drops except in the flash vaporizers and heat exchangers.

8. There are no heat losses except through the heat exchangers.

9. All liquid is discharged at atmospheric pressure.

10. The temperature difference between the water and organic fluid at the inlet and exit of each heat exchanger is equal to the specified pinch-point temperature difference.

11. The temperature difference between the water and organic fluid within the heat exchangers is equal to or greater than the specified pinch-point temperature difference.

12. The organic fluid enters the last-stage heat exchanger at the condensing temperature TR.

Assumption 1 represents ideal operation of the separators with no carry-over of moisture or salt in the steam. Assumption 2 is valid for all practical conditions; all of the steam will be condensed at heat transfer conditions of interest.

Assumption 3 represents the desirable condition of matching temperatures between stages. The energy recoverable from the pressure difference between heat exchangers is small and, according to Assumption 4, is merely dissipated in throttling.

Assumption 5 represents ideal operation of the separators in preventing carry-over of vapor with the liquid. Assumption 6 simplifies the analysis and introduces little error.

Assumptions 7 and 8 simplify the analysis and introduce little error since the effects of pressure drops and heat losses can be included in the power cycle efficiency. Assumption 9 meets usual pressure requirements for liquid discharge to reinjection wells, cooling towers, or surface disposal.

Assumption 10 represents the desirable condition of minimizing the temperature difference between the water and organic fluid. The assumed condition of fixed temperature difference at the inlet and exit of each heat exchanger can always be attained by adjustment of flash vaporization temperatures and organic fluid flow rate. Assumption 11, that the temperature difference inside the heat exchangers does not fall below the temperature difference at the ends, is always true for the vapor heat heat exchangers but may be violated in the liquid heat exchanger, in which case the predicted output power will be slightly too high.

Assumption 12 is true if the organic fluid pump is cooled to TR, and approximately true even for adiabatic pumping because of the low compressibility of the organic fluid at the pumping temperature.

D. Flash Vaporizer

The conditions at the exit of the flash vaporizer are calculated from Eqs. (D-1) - (D-8) of Appendix D by setting the expansion efficiency ETA to zero.

E. Stage Heat Transfer

The flow entering the flash vaporizer (Figure E-1) has temperature T0, total flow rate MTFV, and quality XIN. The latter is equal to the wellhead quality XW for the first stage and is zero for the other stages. The mixture expands to temperature T1. Equations (D-1) - (D-8) give the exit quality XOUT. The vapor flow rate at the exit of the flash vaporizer is

$$MGFV = XOUT\ MTFV \tag{E-1}$$

and the liquid flow rate is

$$MLFV = MTFV - MGFV. \tag{E-2}$$

The flow rate of the vapor leaving the separator is MGFV, and the flow rate of the liquid leaving the heat exchanger of the previous stage is MHX_{n-1}. Thus, the flow rate entering the heat exchanger is

$$MHX = MGFV + MHX_{n-1}. \tag{E-3}$$

The quality of the flow entering the heat exchanger is

$$XHX = \frac{MGFV}{MHX} \tag{E-4}$$

The vapor-phase is condensed at temperature T1, and the liquid is cooled to temperature T2. Since no work is done, the heat transferred to the organic fluid is equal to the enthalpy change. The enthalpy of the vapor entering from the separator is MHX XHX HGB(T1), where HGB(T1) is the enthalpy of vapor in equilibrium with brine at T1. The enthalpy of the water entering from the previous heat exchanger can be evaluated at any convenient state reached in the throttle, such as saturation, since there is no enthalpy change during throttling; thus the entering water enthalpy is MHX (1-XHX) HL(T1), where HL(T1) is the enthalpy of saturated water at T1. Similarly, the enthalpy of the water leaving the heat exchanger at temperature T2 can be evaluated at saturation. Hence the heat transferred to the organic fluid is

$$QGX = MHX\left[(1-XHX)\ HL(T1) + XHX\ HGB(T1) - HL(T2)\right] \quad \text{(E-5)}$$

If the optional last-stage liquid heat exchanger is used (binary process), the additional heat transferred is the enthalpy difference for cooling the brine of flow rate MLFV from T1 to T2.

$$QLX = MLFV\left[HB(T1) - HB(T2)\right] \quad \text{(E-6)}$$

where HB(T) is the enthalpy of brine at T.

Without the liquid heat exchanger the heat transferred is

$$QHX = QGX \quad \text{(E-7)}$$

and with the liquid heat exchanger the heat transferred is

$$QHX = QGX + QLX. \quad \text{(E-8)}$$

F. Organic Fluid Flow Rate

The total heat transferred to the organic fluid is

$$QSUM = \sum_{1}^{NSTG} QHX \quad \text{(E-9)}$$

The initial temperature of the organic fluid, by Assumption 12, is the condensing temperature TR. The final temperature, by Assumption 10, is

$$TFMAX = TX(2) - TPINCH \quad \text{(E-10)}$$

where TX(2) is the temperature in the first-stage separator.

If HF(T) is the enthalpy of the organic fluid at temperature T and heat exchanger pressure, the enthalpy change of the organic fluid per unit flow rate is HF(TFMAX) - HF(TR). Since the organic fluid enthalpy change, by Assumption 8, must equal the total heat exchanger heat transfer QSUM, the flow rate of the organic fluid is

$$MF = \frac{QSUM}{HF(TMAX) - HF(TR)} \quad \text{(E-11)}$$

G. Organic Fluid Temperatures

The exit temperature of the organic fluid from the first-stage heat exchanger is TF2 = TFMAX from Eq. (E-10), and the exit enthalpy per unit mass is

$$HF2 = HF(TF2). \quad \text{(E-12)}$$

The inlet enthalpy of the organic fluid is equal to the exit enthalpy minus the heat transferred. Thus, the inlet enthalpy per unit mass is

$$HF1 = HF2 - \frac{QHX}{MF} \tag{E-13}$$

The corresponding inlet temperature is

$$TF1 = TF(HF1) \tag{E-14}$$

where TF(HF1) is the temperature of the organic fluid at enthalpy HF1 and heat exchanger pressure, obtained from thermodynamic property data.

For each succeeding stage the exit temperature TF2 is set equal to the inlet temperature of the preceding stage and the procedure of Eqs. (E-12) - (E-14) repeated to find the inlet temperature TF1. The TF1 value found for the last stage will be TR since the organic fluid flow rate was chosen for an overall temperature change from TFMAX to TR.

H. Flash Vaporization Temperatures

According to Assumption 11 the pinch points are located at the inlets and exits of the heat exchangers. Thus, the flash vaporization temperatures are

$$\begin{aligned} TX(3) &= TF1_1 + TPINCH \\ TX(4) &= TF1_2 + TPINCH \\ &\vdots \\ TX(NSTG+1) &= TF1_{NSTG-1} + TPINCH \end{aligned} \tag{E-15}$$

The procedure for finding the flash vaporization temperatures is to make an initial guess, calculate the heat exchanger heat transfer rates from Eqs. (E-1)-(E-8), the organic fluid flow rate from Eqs. (E-9)- (E-11), the organic fluid temperatures from Eqs. (E-12)-(E-14), and find a new set of flash vaporization temperatures from Eq. (E-15). Repeating the procedure gives convergence of the temperatures to the correct values in about ten iterations.

The initial temperature guess is based on equal temperature drops across the heat exchangers starting with TX(2).

$$DT = \frac{TX(2) - (TR + TPINCH)}{NSTG} \tag{E-16}$$

The temperatures are set to TX(3) = TW-DT, TX(4) = TW-2DT, and so on. The inlet temperature TX(2) to the first heat exchanger is either specified or

optimized for maximum output power.

I. Liquid Discharge Pumping Power

The pressure of the brine in the last-stage separator is PD = PSATB(T1), the saturation pressure of brine at the last-stage separator temperature. The pressure of the water leaving the last-stage heat exchanger is the same. If PD is below atmospheric pressure it is necessary, according to Assumption 9, to pump the liquid to atmospheric pressure. Except for the small difference in brine density, the pumping power is the same whether the brine is discarded from the separator or cooled in the optional last-stage liquid heat exchanger. Using T1 for the brine temperature in either case, the pumping power is

$$PLP = \left[\frac{MHX}{RHOL(T2)} + \frac{MLFV}{RHOB(T1)}\right] \frac{PATM - PD}{ETAP} \qquad \text{(E-17)}$$

where RHOL(T2) and RHOB(T1) are the densities of water and brine, respectively, at T2 and T1, and ETAP is the pump efficiency.

V. Output Power

The maximum work available from a fluid between states 1 and 2, with a heat sink available at absolute temperature TRA, is $W=H_1 - H_2 - TRA(S_1-S_2)$, where H_1 and S_1 are the enthalpy and entropy at state 1 and H_2 and S_2 are the enthalpy and entropy at state 2. (This relation comes from considering a series of carnot-cycle engines accepting heat from the fluid as it cools and rejecting heat at TRA).

For the organic fluid the initial state is TFMAX and the final state is TR, both at the heat exchanger pressure. The heat sink temperature, for TR in ^{o}C, is

$$TRA = TR + 273.18 \qquad \text{(E-18)}$$

Thus, the available power in the organic fluid at flow rate MF is

$$PB = HF(TFMAX) - HF(TR) - TRA\left[SF(TFMAX) - SF(TR)\right] \qquad \text{(E-19)}$$

where SF(T) is the entropy of the organic fluid at temperature T and heat exchanger pressure.

The output power of the organic cycle is a specified fraction ETAB of the available power in the organic fluid. The net output power of the process, after deducting any pumping power for returning the brine to atmospheric pressure, is

$$PNET = ETAB\ PB - PLP. \qquad \text{(E-20)}$$

K. Power Recovery Fraction and Efficiency

The available work AVAIL from unit mass of the wellhead mixture (the specific available wellhead power) is given by Eqs. (D-1) - (D-8) of Appendix D for X1 = XW, T1 = TW, T2 = TR, ETA = 1, and M = 1. The power recovery fraction is the ratio of net output power to wellhead available power

$$PFR = \frac{PNET}{AVAIL} \tag{E-21}$$

The quantity that can be considered the equivalent heat input to the process is the difference between the enthalpy of the wellhead mixture and the enthalpy of brine at the condensing temperature TR, since that is the heat that would have to be supplied to heat the brine from TR to the wellhead conditions. That heat is

$$QIN = (1 - XW)\ HB(TW) + XW\ HGB(TW) - HB(TR) \tag{E-22}$$

The corresponding cycle efficiency of the process is

$$ETAC = \frac{PNET}{QIN} \tag{E-23}$$

L. Condenser Heat Rejection

The heat rejected by the organic cycle condenser is just the difference between the heat input to the organic cycle and the power output.

$$QC = QSUM - ETAB\ PB \tag{E-24}$$

M. Value of Factor ETAB for a Rankine Cycle Using Isobutane

The factor ETAB in Eq. (E-20) is unity if the organic turbine and pump have efficiencies of 1.0 and all of the heat is rejected at the condensing temperature TR. The value of ETAB for an actual organic Rankine cycle with turbine and pump efficiencies of 0.8 and isobutane as the working fluid will be calculated to determine a realistic value of ETAB.

Figure E-3 shows a Rankine cycle using isobutane[E-1] as the working fluid. Liquid isobutane leaves the condenser at point 1 at condensing temperature TR = 45°C and saturation pressure 0.60 MPa (87 psi). In the pump, an isentropic compression to 4.14 MPa (600 psi) requires an enthalpy increase of 5.9 J/g. At a pump efficiency of 0.8 the work supplied to the pump is 5.9/0.8 = 7.4 J/g. The pump exit temperature (point 2) for adiabatic flow is 47°C.

The isobutane is then heated to 180°C (point 3), requiring a heat input of 524.1 J/g. In the turbine, an isentropic expansion to the condensing pressure of 0.60 MPa yields an enthalpy change of 99.8 J/g. At a turbine efficiency of 0.8 the work delivered by the turbine is 99.8 x 0.8 = 79.8 J/g. The turbine exit temperature (point 4) is 115°C.

The superheated isobutane is cooled to saturation at 45°C (point 5), rejecting 142.6 J/g of heat. The isobutane is then condensed, rejecting 309.1 J/g of heat and completing the cycle.

The net work of the cycle is the difference between the turbine work and pump work, 79.8 - 7.4 = 72.4 J/g. The available work between 45°C and 180°C at the heat exchanger pressure, calculated from Eq. (E-19), is 101.7 J/g. Thus, the value of ETAB for this cycle is 72.4/101.7 = 0.71, showing that 0.7 is a realistic value for ETAB.

N. Computer Program

A Fortran IV computer program incorporating the equations derived in this Appendix is presented on the following pages. Results for a test case are also given. The program contains enthalpy and entropy data for isobutane at 4.14 MPa (600 psi) from Reference E-1.

E-1. Starling, K. E., "Fluid Thermodynamic Properties for Light Petroleum Systems", Gulf Publishing Company, Houston, Texas, 1973, pp. 194-205.

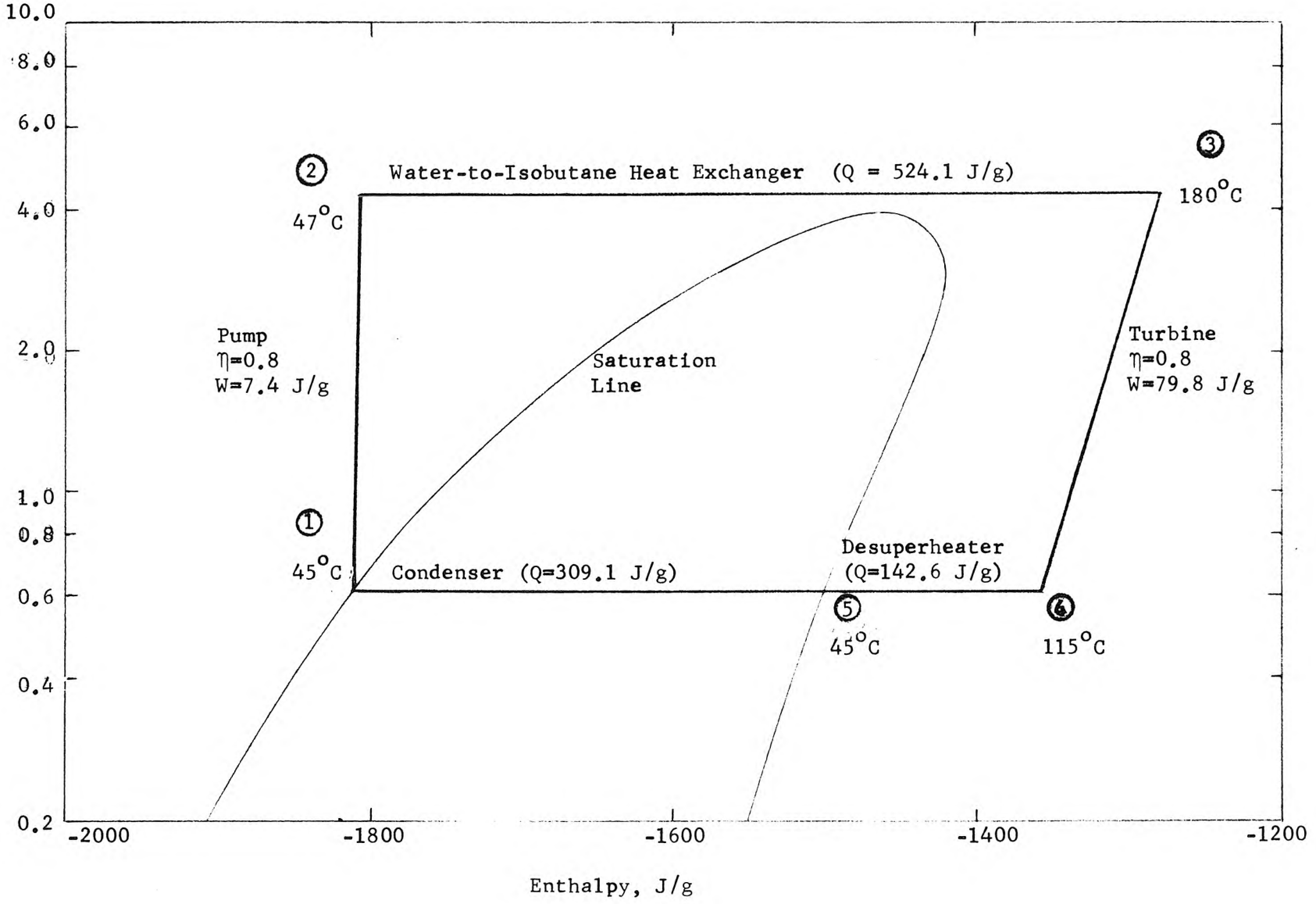

Fig. E-3 Rankine cycle between 180°C and 45°C using isobutane as the working fluid

```
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C  * * * * * * * *  BINARY AND FLASH BINARY PROCESSES  * * * * * * * * *
C
C  * * * * * * * * * * D. G. ELLIOTT, MAY 15, 1975 * * * * * * * * * * *
C
C  * * * * * * * * * * LAST REVISION AUG. 18, 1975 * * * * * * * * * * *
C
C  CALCULATES THE POWER AVAILABLE FROM AN ORGANIC WORKING FLUID CON-
C  DENSING AT SPECIFIED REJECTION TEMPERATURE TR AND HEATED BY 1 KG/S
C  OF BRINE AT WELLHEAD TEMPERATURE TW AND QUALITY XW. THE HEATING IS
C  IN NSTG STAGES OF FLASH VAPORIZATION WITH HEAT TRANSFER FROM THE
C  FLASHED STEAM AND THE CONDENSATE CARRIED OVER FROM THE PREVIOUS
C  STAGES, PLUS, OPTIONALLY, HEAT TRANSFER FROM THE LIQUID LEAVING
C  THE LAST-STAGE.
C
C  THE ORGANIC FLUID FOR WHICH DATA IS INCLUDED IN THE PROGRAM IS
C  ISOBUTANE AT 600 PSI HEAT EXCHANGER PRESSURE. OTHER FLUIDS OR
C  PRESSURES CAN BE USED BY CHANGING THE HT AND ST DATA ARRAYS.
C
C  * * * * * * * * EQUATION NUMBERS REFER TO APPENDIX E  * * * * * * * *
C
C  INPUT DATA
C  ----------
C
C  CONC     BRINE CONCENTRATION FRACTION (DENVER RESEARCH INSTITUTE MIX)
C  TW       WELLHEAD TEMPERATURE, DEG C
C  XW       WELLHEAD QUALITY
C  NSTG     NUMBER OF STAGES
C  TR       ORGANIC FLUID CONDENSING TEMPERATURE, DEG C
C  TPINCH   PINCH POINT TEMPERATURE DIFFERENCE, DEG C
C  ETAB     RATIO OF ORGANIC CYCLE OUTPUT TO ORGANIC FLUID AVAILABLE PWR
C  ETAP     EFFICIENCY OF PUMPS TO RETURN LIQUID TO ATMOSPHERIC PRESSURE
C  NINPT    '1' FOR ECHO OF INPUTS
C  THX1      FIRST-STAGE SEPARATOR TEMPERATURE (NOT REQUIRED IF NSRCH=1)
C  NSRCH    '1' FOR OPTIMIZATION OF FIRST-STAGE SEPARATOR TEMPERATURE
C  DTSRCH    LARGEST TEMPERATURE DIFFERENCE BELOW TW FOR THX1 SEARCH
C  NLX      '1' FOR LIQUID HEAT EXCHANGER IN LAST STAGE
C
C  OUTPUTS (FOR 1 KG/S WELLHEAD FLOW RATE)
C  ---------------------------------------
C
C  STAGE OUTPUTS:
C
C  STG      STAGE NUMBER
C  XOUT     QUALITY AT EXIT OF FLASH VAPORIZER
C  MHX      HEAT EXCHANGER BRINE-SIDE FLOW RATE, KG/S
C  T0       TEMPERATURE AT INLET OF FLASH VAPORIZER, DEG C
C  T1       INLET TEMP OF THE HEAT EXCHANGER ON THE WATER SIDE, DEG C
C  T2       EXIT TEMP OF THE HEAT EXCHANGER ON THE WATER SIDE, DEG C
C  QHX      HEAT EXCHANGER HEAT TRANSFER RATE, KW
```

```
C   TF2      EXIT TEMP OF THE HEAT EXCHANGER ON THE ORGANIC SIDE, DEG C
C   TF1      INLET TEMP OF THE HEAT EXCHANGER ON THE ORGANIC SIDE, DEG F
C
C   TOTALS:
C
C   MF         ORGANIC FLUID FLOW RATE, KG/S
C   PLP        PUMP POWER TO RETURN LIQUID TO ATMOSPHERIC PRESS, KW
C   PNET       NET OUTPUT POWER (ORGANIC FLUID AVAIL PWR)*ETAB-PLP, KW
C   QC         CONDENSER HEAT REJECTION, KW
C   QIN        ENTHALPY OF WELLHEAD FLOW RELATIVE TO LIQUID AT TR, KW
C   AVAIL      AVAILABLE WELLHEAD POWER, KW
C   ETAC       EQUIVALENT CYCLE EFFICIENCY, PNET/QIN
C   P/AVAIL    POWER RECOVERY FRACTION, PNET/AVAIL
C
      PARAMETER MXT=300, MXSTG=6, MXTX=MXSTG+2
C
      INTEGER T0,T1,T2,TW,TR,TMAX,TMIN,TT,TSAVE,DTSRCH,THX1,TX(MXTX)
      IMPLICIT REAL (M)
      REAL PSAT(MXT),HL(MXT),HG(MXT),SL(MXT),SG(MXT),RHOL(MXT),
     &     RHOG(MXT),PSATB(MXT),HB(MXT),HGB(MXT),SB(MXT),SGB(MXT),
     &     RHOB(MXT),RHOGB(MXT),TTAB(23),HT(23),ST(23),TFIN(MXTX)
C
C     EQUIVALENCE (PSAT,PSATB),(HL,HB),(HG,HGB),(SL,SB),(SG,SGB),
C    &            (RHOL,RHOB),(RHOG,RHOGB) @ FOR PURE WATER CASES
C
      NAMELIST /IN/CONC,TW,XW,NSTG,TR,TPINCH,ETAB,ETAP,NINPT,THX1,
     &          NSRCH,DTSRCH,NLX
C
C ***** ISOBUTANE H AND S FOR T FROM 40 F TO 480 F AT 600 PSI *****
C  SOURCE: STARLING, 'FLUID THERMODYNAMIC PROPERTIES OF LIGHT
C          PETROLEUM SYSTEMS'
C
      DATA HT/-818.17, -807.28, -796.12, -784.65, -772.82, -760.58,
     &        -747.85, -734.51, -720.40, -705.25, -688.54, -669.09,
     &        -643.17, -599.01, -577.62, -561.10, -546.01, -531.55,
     &        -517.39, -503.37, -489.37, -475.35, -461.26/
      DATA ST/0.8530, 0.8744, 0.8956, 0.9166, 0.9374, 0.9583, 0.9793,
     &        1.0006, 1.0224, 1.0450, 1.0693, 1.0968, 1.1323, 1.1912,
     &        1.2190, 1.2398, 1.2584, 1.2757, 1.2923, 1.3082, 1.3237,
     &        1.3389, 1.3538/
          DO 5 K=1,23
5         TTAB(K)=20+20*K @ TABLE TEMPERATURES FOR HT AND ST, DEG F
C
C  ***** PURE WATER THERMODYNAMIC PROPERTIES *****
C
      CALL H2OPRP(PSAT, HL, HG, SL, SG, RHOL, RHOG,
     &            PSATB, HB, HGB, SB, SGB, RHOB, RHOGB, MXT)
C
C  ***** DEFAULT INPUTS *****
C
      NSTG=1 @ ONE STAGE
      TR=45 @ 45 DEG C ORGANIC FLUID CONDENSING TEMPERATURE
```

```
      TPINCH=10 @ 10 DEG C PINCH POINT TEMPERATURE DIFFERENCE
      ETAB=0.7 @RATIO OF ORGANIC CYCLE OUTPUT TO ORGANIC FLUID AVAIL PWR
      ETAP=0.7 @ LIQUID PUMP EFFICIENCY
      DTSRCH=20 @ SEARCH OVER 20 DEG C RANGE BELOW TW FOR THX1
C
C  ***** INPUTS AND CONSTANTS *****
C
10    PRINT 15
15    FORMAT (//' ENTER')
      READ (5,IN,ERR=10,END=95) @ READ INPUTS
      IF (TW.GT.MXT) PRINT 20
20    FORMAT (' MXT TOO SMALL')
      IF (NSTG.GT.MXSTG) PRINT 21
21    FORMAT (' MXSTG TOO SMALL')
      IF (TW.GT.MXT.OR.NSTG.GT.MXSTG) GO TO 10
      TX(1)=TW @ SET FIRST TEMP IN ARRAY TO WELLHEAD TEMP
      TX(NSTG+2)=TR+TPINCH @SET LAST TEMP TO CONDENSING TEMP PLUS TPINCH
      IF (NINPT.EQ.1) WRITE (6,IN) @ ECHO INPUTS
      PATM=1.013E5 @ ATMOSPHERIC PRESSURE, PA
      NITER=9 @ NO. OF ITERATIONS FOR RESETTING TX'S (USE ODD NUMBER)
      TRR=TR @ REJECTION TEMPERATURE, DEG C, REAL
      TRA=TRR+273.18 @ ABSOLUTE REJECTION TEMP, DEG K (18)
      TMAX=THX1 @ FOR NON-OPTIMIZATION MODE
      TMIN=THX1 @ FOR NON-OPTIMIZATION MODE
      PMAX=0.0 @ INITIALIZE MAX POWER FOR SEARCH
C
C  ***** BRINE THERMODYNAMIC PROPERTIES *****
C
      OLDCON=-1 @ TO ASSURE CALL ON FIRST PASS
      IF (CONC.NE.OLDCON) CALL BRNPRP(CONC) @ LOAD BRINE PROPERTIES
      OLDCON=CONC @ TO PREVENT REPEAT CALL IF CONC NOT CHANGED
C
C  ***** AVAILABLE WELLHEAD POWER *****
C
      CALL EXPAND (1.0, XW, SB(TW), SGB(TW), SB(TR), SGB(TR), HB(TW),
     &  HGB(TW), HB(TR), HGB(TR), 1.0, XR, AVAIL) @ AVAIL PWR (D1-D8)
      IF (NSRCH.EQ.0) GO TO 25 @ NO OPTIMIZATION
C
C  **** OPTIONAL OPTIMIZATION OF FIRST-STAGE SEPARATOR TEMPERATURE *****
C
      TMAX=TW @ MAX SEARCH TEMP
      IF (XW.EQ.0) TMAX=TW-1 @ MAX SEARCH TEMP WITH ZERO QUAL
      TMIN=TMAX-DTSRCH @ LOWEST SEARCH TEMP
      NX=1 @ TO SUPPRESS PRINTOUT UNTIL COMPLETION OF SEARCH
C
25    DO 75 TT=TMAX,TMIN,-1 @ SCAN TX(2) FROM TMAX TO TMIN
      TX(2)=TT
      DT=(TX(2)-TR-TPINCH)/NSTG @ GUESS AT TEMP DIFF PER STAGE (16)
      IF (NSTG.LT.2) GO TO 35 @ NO INTERMEDIATE TEMPERATURES TO SET
           DO 30 K=2,NSTG @ LOAD STAGE TEMPERATURE GUESSES
30         TX(K+1)=TW-DT*(K-1) @ EQUAL TEMPERATURE STEPS
C
```

```
C  * * * * * * *  S T A G E  C A L C U L A T I O N S  * * * * * * *
C
C  ***** ITERATION FOR PINCH POINTS *****
C
35    TFMAX=TX(2)-TPINCH @ ORGANIC FLUID MAXIMUM TEMPERATURE (10)
      DO 65 KK=1,NITER @ SET SEPARATOR TEMPS TO ORGANIC TEMPS +TPINCH
      IF (NSTG.LT.2) GO TO 45 @ NO INTERMEDIATE TX'S TO SET
           DO 40 K=2,NSTG
40         IF (KK.GT.1) TX(K+1)=TFIN(K-1)+TPINCH+0.5 @ RESET TX'S (15)
45    DO 65 J=1,2 @ FIND ORGANIC FLUID FLOW RATE (J=1) AND POWERS (J=2)
      MLFV=1.0 @ WILL BE FLASH VAPORIZER INLET FLOW IN FIRST STAGE
      MHX=0.0 @ NO PREVIOUS FLOW TO FIRST STAGE HEAT EXCHANGER
      TF1=TFMAX @ WILL BE ORGANIC FLUID EXIT TEMP FROM FIRST STAGE
      QSUM=0.0 @ INITIALIZE TOTAL HEAT TRANSFER SUM TO ZERO
C
           DO 60 K=1,NSTG @ CALC FLOWS AND HEAT TRANSFER IN EACH STAGE
           T0=TX(K) @ INLET TEMPERATURE TO FLASH VAPORIZER
           T1=TX(K+1) @ EXIT TEMP OF FLASH VAP AND HEAT-X INLET TEMP
           T2=TX(K+2) @ HEAT EXCHANGER EXIT TEMP
C
C  ***** FLASH VAPORIZER *****
C
           XIN=0 @ QUALITY FOR ALL-LIQUID INLET
           IF (K.EQ.1) XIN=XW @ USE WELLHEAD QUALITY FOR FIRST STAGE
           MTFV=MLFV @ TOTAL INLET FLOW =PREV STAGE LIQUID OUT FLOW
           CALL EXPAND (MTFV, XIN, SB(T0), SGB(T0), SB(T1), SGB(T1),
     &     HB(T0), HGB(T0), HB(T1), HGB(T1), 0.0, XOUT, PWR) @ (D1-D8)
           MGFV=XOUT*MTFV @ VAPOR FLOW LEAVING FLASH VAPORIZER (1)
           MLFV=MTFV-MGFV @ LIQUID FLOW LEAVING FLASH VAPORIZER (2)
C
C  ***** HEAT EXCHANGER *****
C
           MHX=MHX + MGFV @ FLOW RATE IN HEAT EXCHANGER (3)
           XHX=MGFV/MHX @ QUALITY AT ENTRANCE OF HEAT EXCHANGER (4)
           QGX=MHX*((1-XHX)*HL(T1) + XHX*HGB(T1) - HL(T2)) @VAP HT X (5)
           QLX=MLFV*(HB(T1)-HB(T2)) @ LIQUID HT X HEAT AVAILABLE (6)
           QHX=QGX @ STAGE HEAT TRANSFER WITHOUT LIQUID HT X (7)
           IF (K.EQ.NSTG.AND.NLX.EQ.1) QHX=QHX+QLX @ WITH LIQ HT X (8)
           QSUM=QSUM+QHX @ CUMULATIVE HEAT TRANSFER TO ORGANIC FLUID (9)
           IF (J.EQ.1) GO TO 60 @ JUST FINDING QSUM ON FIRST PASS
C
           TF2=TF1 @ ORGANIC FLUID EXIT TEMP=PREVIOUS STAGE INLET TEMP
           HF2=HF(TF2) @ ENTHALPY OF ORG FLUID AT STAGE EXIT TEMP (12)
           HF1=HF2 - QHX/MF @ ENTHALPY OF ENTERING ORGANIC FLUID (13)
           TF1=TF(HF1) @ TEMPERATURE OF ENTERING ORGANIC FLUID (14)
           IF (K.EQ.NSTG) TF1=TRR @ LAST STAGE
           TFIN(K)=TF1 @ SAVE ORGANIC FLUID INLET TEMP
C  ***** PRINTOUT OF STAGE QUANTITIES *****
C
           IF (KK.EQ.NITER.AND.K.EQ.1.AND.NX.EQ.0) PRINT 50 @ HEADING
50          FORMAT (//
     &     ' STG  XOUT    MHX    T0    T1    T2    QHX     TF2     TF1'/)
```

```
            IF (KK.EQ.NITER.AND.NX.EQ.0) PRINT 55,K,XOUT,MHX,T0,T1,
     &      T2,QHX,TF2,TF1 @ STAGE PRINTOUT
55          FORMAT (I3, F7.3, F6.3, 3(I5), -3PF7.1, 2(0PF7.1))
60          CONTINUE
65    MF=QSUM/(HF(TFMAX)-HF(TRR)) @ ORGANIC FLUID FLOW RATE (11)
C
C  ***** PUMP POWER *****
C
      PD=PSATB(T1) @ LAST-STAGE SEPARATOR PRESSURE
      PLP=0.0 @ PUMP PWR IF CONDENSING PRESSURE IS ABOVE ATMOSPHERIC
      IF (PD.LT.PATM)
     &  PLP=(MHX/RHOL(T2)+MLFV/RHOB(T1))*(PATM-PD)/ETAP @ PUMP PWR (17)
C
C  ***** POWER RECOVERY FRACTION *****
C
      PB=MF*(HF(TFMAX)-HF(TRR)-TRA*(SF(TFMAX)-SF(TRR))) @ AVAIL PWR (19)
      PNET=ETAB*PB-PLP @ NET OUTPUT POWER (20)
      PRF=PNET/AVAIL @ POWER RECOVERY FRACTION (21)
      IF (NX.EQ.0) GO TO 85 @ TX(2) SEARCH SCAN COMPLETED
C
C  ***** COMPLETION OF TEMPERATURE OPTIMIZATION LOOP *****
C
      IF (NSRCH.EQ.2) PRINT 70,TT,PNET @ TO MONITOR SEARCH
70    FORMAT ()
      IF (PNET.GT.PMAX) TSAVE=TT @ THIS WILL BE TX(2) FOR HIGHEST PNET
75    IF (PNET.GT.PMAX) PMAX=PNET
C
      TMAX=TSAVE @ SET DO-75 LOOP FOR ONCE THRU AT BEST TEMP TX(2) VALUE
      IF (TSAVE.EQ.TMIN) PRINT 80,TSAVE
80    FORMAT (' STILL IMPROVING AT TX(2)=',I4/)
      TMIN=TMAX
      NX=0 @ SET FLAG FOR LAST TIME THROUGH
      GO TO 25 @ LAST TIME THROUGH
C
C  ***** CONDENSER HEAT REJECTION AND EQUIVALENT EFFICIENCY *****
C
85    QC=QSUM-ETAB*PB @ ORGANIC CYCLE CONDENSER HEAT REJECTION (24)
      QIN=(1-XW)*HB(TW) + XW*HGB(TW) - HB(TR) @ EQUIV HEAT INPUT (22)
      ETAC=PNET/QIN @ EQUIVALENT CYCLE EFFICIENCY (35)
C
C  ***** PRINTOUT OF TOTALS *****
C
      PRINT 90 @ HEADING
90    FORMAT (//4X
     &'MF    PLP      PNET       QC       QIN      AVAIL    ETAC  P/AVAIL'/)
      PRINT 91,MF,PLP,PNET,QC,QIN,AVAIL,ETAC,PRF @ PRINT TOTALS
91    FORMAT(F6.2,-3PF6.2, 4(F9.2), 2(0PF7.4))
C
      GO TO 10 @ ASK FOR NEW INPUTS
C
95    STOP
C
```

```
C  * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * * *
C
C  SUBROUTINE EXPAND CALCULATES THE EXIT QUALITY AND POWER OUTPUT OF A
C  TWO PHASE EXPANSION PROCESS GIVEN THE FLOW RATE, INLET QUALITY,
C  INLET AND EXIT THERMODYNAMIC PROPERTIES, AND EFFICIENCY.
C
C  ***** EQUATION NUMBER REFER TO APPENDIX D *****
C
      SUBROUTINE EXPAND (M, X1, SL1, SG1, SL2, SG2,
     &                   HL1, HG1, HL2, HG2, ETA, X2, P)
C
      S1=(1-X1)*SL1 + X1*SG1 @ INLET MIXTURE ENTROPY (1)
      H1=(1-X1)*HL1 + X1*HG1 @ INLET MIXTURE ENTHALPY (2)
      X2I=(S1-SL2)/(SG2-SL2) @ ISENTROPIC EXIT QUALITY (3)
      H2I=(1-X2I)*HL2 + X2I*HG2 @ ISENTROPIC EXIT ENTHALPY (4)
      DH=ETA*(H1-H2I) @ ACTUAL ENTHALPY CHANGE (5)
      H2=H1-DH @ EXIT ENTHALPY (6)
      X2=(H2-HL2)/(HG2-HL2) @ EXIT QUALITY (7)
      P=M*DH @ EXPANDER POWER OUTPUT (8)
C
      FUNCTION HF(T) @ ORGANIC FLUID ENTHALPY AS A FUNCTION OF TEMP
      X=1.8*T+32 @ DEG F
      CALL SLUP(X,Y,YDOT,TTAB,HT,23,2) @ 2'ND-ORDER INTERP FOR ENTHALPY
      HF=Y*2326.0 @ ENTHALPY IN J/KG
      RETURN
C
      FUNCTION SF(T) @ ORGANIC FLUID ENTROPY AS A FUNCTION OF TEMP
      X=1.8*T+32 @ DEG F
      CALL SLUP(X,Y,YDOT,TTAB,ST,23,2) @ 2'ND ORDER INTERP FOR ENTROPY
      SF=Y*4186.8 @ ENTROPY IN J/KG*C
      RETURN
C
      FUNCTION TF(H) @ ORGANIC FLUID TEMP AS A FUNCTION OF ENTHALPY
      X=H/2326 @ ENTHALPY IN BTU/LB
      CALL SLUP(X,Y,YDOT,HT,TTAB,23,2) @ 2'ND ORDER INTERP FOR TEMP
      TF=(Y-32)/1.8 @ TEMP, DEG C
      RETURN
C
      END
```

```
$IN CONC=0.1, TW=210, XW=0.05, NSTG=3, NINPT=1, THX1=196 $END


ENTER
$IN
CONC     =      .10000000E+00
TW       =              +210
XW       =      .50000000E-01
NSTG     =                +3
TR       =               +45
TPINCH   =      .10000000E+02
ETAB     =      .70000000E+00
ETAP     =      .70000000E+00
NINPT    =                +1
THX1     =              +196
NSRCH    =                +0
DTSRCH   =               +20
NLX      =                +0
$END
```

STG	XOUT	MHX	TO	T1	T2	QHX	TF2	TF1
1	.071	.071	210	196	151	153.6	186.0	141.4
2	.067	.133	196	151	114	152.4	141.4	104.0
3	.053	.179	151	114	55	147.0	104.0	45.0

MF	PLP	PNET	QC	QIN	AVAIL	ETAC	P/AVAIL
.83	.00	59.83	393.20	657.50	144.09	.0910	.4152

#25741